U.S. DEPARTMENT OF COMMERCE
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PB-264 536

Comparison of Flue Gas Desulfurization Coal Liquefaction for Use at Coal-Fired Power Plants

M. W. Kellogg Co, Houston, Tex. Research & Engineering Dev.

Prepared for

Environmental Protection Agency, Research Triangle Park, N C Office of Air Quality Planning and Standards

Apr 75

PB 264 536

EPA-450/3-75-047

April 1975

COMPARISON OF FLUE GAS DESULFURIZATION, COAL LIQUEFACTION, AND COAL GASIFICATION FOR USE AT COAL-FIRED POWER PLANTS

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SPRINGFIELD, VA. 22161

U.S. ENVIRONMENTAL PROTECTION AGENCY
Office of Air and Waste Management
Office of Air Quality Planning and Standards
Research Triangle Park, North Carolina 27711

TECHNICAL REPORT DATA

(Please read Instructions on the reverse before completing)

1. REPORT NO. EPA-450/3-75-047	2.	3. RECIPIENT'S ACCESSION NO.
4. TITLE AND SUBTITLE Comparison of Flue Gas Desulfurization, Coal Liquefaction, and Coal Gasification for Use at Coal-Fired Power Plants		5. REPORT DATE
		6. PERFORMING ORGANIZATION CODE
7. AUTHOR(S)		8. PERFORMING ORGANIZATION REPORT NO.
9. PERFORMING ORGANIZATION NAME AND ADDRESS The M. W. Kellogg Company Research and Engineering Development 1300 Three Greenway Plaza East Houston, Texas 77046		10. PROGRAM ELEMENT NO.
		11. CONTRACT/GRANT NO. No. 68-02-1308
12. SPONSORING AGENCY NAME AND ADDRESS U. S. Environmental Protection Agency Research Triangle Park, North Carolina 27711		13. TYPE OF REPORT AND PERIOD COVERED Final Report
		14. SPONSORING AGENCY CODE

15. SUPPLEMENTARY NOTES

16. ABSTRACT

The report presents a technical and economic comparison of the use of flue gas desulfurization, coal liquefaction, and coal gasification as a means of preventing SO_2 emissions at coal-fired power plants. The report assesses the status of technology, process complexity, process flexibility, environmental effects, installation difficulties, energy conversion efficiency, manpower requirements, and economics of each approach to controlling SO_2 as it would be applied to electric power plants. Three different flue gas desulfurization systems were evaluated as well as one coal liquefaction and one coal gasification processes. Two power plant cases are evaluated, an existing 500 MW plant operating at 60 percent load factor and a new 1,000 MW plant operating at 80 percent load factor.

17. KEY WORDS AND DOCUMENT ANALYSIS

a. DESCRIPTORS	b. IDENTIFIERS/OPEN ENDED TERMS	c. COSATI Field/Group
Air Pollution Chemical Reaction Gasification Liquefaction Desulfurization Economic Analysis Design Sulfur Dioxide Limestone Coal Sulfur Calcium Oxides Combustion Products Flue Gases	Air Pollution Control Electric Power Plants Boilers	13B

18. DISTRIBUTION STATEMENT Unlimited	19. SECURITY CLASS *(This Report)* Unclassified	21. NO. OF PAGES 158
	20. SECURITY CLASS *(This page)* Unclassified	22. PRICE

EPA Form 2220-1 (9-73)

EPA-450/3-75-047

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Office of Air and Waste Management
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Research Triangle Park, North Carolina 27711

April 1975

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1. INTRODUCTION

The work reported herein is a comparison of three different desulfurization techniques:

1) flue gas desulfurization
2) solvent refined coal
3) coal gasification to produce low Btu gas

The study was performed for the Environmental Protection Agency under Task 34, Contract No. 68-02-1308.

In order to meet sulfur oxide emissions standards, combustion sources which normally burn high sulfur fuels can be controlled by removing sulfur in one of three ways:

1) prior to combustion
2) during combustion
3) after combustion

Coal gasification and solvent refined coal represent two methods of pre-combustion sulfur control, while flue gas desulfurization is, of course, a post-combustion control method. Although there are a variety of processes under development which remove sulfur during combustion, none were included in this study.

The overall objective of this task was to make a technical and economic comparison of flue gas desulfurization, solvent refined coal, and coal gasification (low Btu gas). It was a basic premise of this task to confine the study to an investigation of these processes as applied to conventional steam-electric power plants. Therefore, low Btu gas for use as fuel in a combined steam and gas turbine cycle was not considered. The latter is a promising technology for base-load plants in the future. However, widespread commercialization is dependent upon successful gas turbine development to allow the turbines to operate at temperatures high enough to achieve better cycle efficiencies than can be obtained in a conventional power plant.

Three flue gas desulfurization systems were included as being representative of the field. These are:

1) the wet limestone (or lime) process
2) the Wellman-Lord/Allied Chemical process
3) the Cat-Ox (Monsanto) process

The solvent refined coal process is based on the Pittsburg and Midway Coal Mining Company flow sheet, while low Btu gas is based on Lurgi pressure gasification.

Each process or technology was reviewed to obtain the following information:

1) process complexity
2) process flexibility
3) status of technology
4) environmental effects
5) installation difficulties
6) energy conversion efficiency
7) manpower requirements
8) economics

The results of this study have been quantified where possible and comparisons between the different technologies have been made based on the available information.

2. SUMMARY AND CONCLUSIONS

This study evaluates flue gas desulfurization, solvent refined coal, and low Btu gas as applied to two different conventional power plants. The first is an existing 500 MW plant operating at 60% load factor, and the second is a new 1000 MW plant operating at 80% load factor. Sulfur recovery efficiencies of 90% have been used for the flue gas desulfurization processes and low Btu gas, giving an overall SO_2 emmission rate of about 0.6 lbs SO_2/MMBtu. For solvent refined coal (SRC), it has been assumed that the SRC can typically be desulfurized to about 1.0% sulfur. This would produce an SO_2 emmission rate from the power plant of approximately 1.2 lbs SO_2/MM Btu. The combined emmissions from the power plant and the SRC plant would be slightly higher.

Of necessity, flue gas desulfurization units and low Btu gas plants must be sized in relation to the power plants which they serve. However, some flexibility in size is possible with solvent refined coal plants. Since the product is easily stored and shipped, these plants need not be integrated with a single power plant, but could serve several power plants within an area. For this study, two different size solvent refined coal plants have been considered. The first is sized to produce fuel corresponding to the power plant fuel consumption. The second is assumed to be four times this size. These plants are identified in subsequent tables according to the equivalent power production from the solvent refined coal product. Thus a 500/500 unit would correspond to a 500 MW power plant being served by a 500 MW (equivalent power) solvent refined coal plant. A 1000/4000 unit would represent a 1000 MW power plant receiving fuel from a 4000 MW (equivalent power) solvent refined coal plant.

The total production costs summarized here are the total operating costs for each process. These include all direct and indirect costs plus depreciation, interest on debt, return on equity, taxes and insurance. Additionally, the cost of coal has been included for each process. This permits a direct comparison to be made between flue gas desulfurization and the other control methods. Costs are shown in cents per million Btu of heat input to the boiler.

2.1 500 MW Existing Plant (60% Load Factor)

Total Production Costs - ¢/MM Btu

	Flue Gas Desulfurization			Solvent Refined Coal (SRC)		Low Btu Gas
Coal Cost $/T	Wet Limestone	Wellman-Lord /Allied	Cat-Ox	500/500	500/2000	
5	57	63	67	117	80	108
10	78	83	88	145	108	143
15	98	104	108	174	137	177

Flue gas desulfurization appears to be superior to the use of SRC or low Btu gasification. The wet limestone process seems to be the least costly scrubbing process followed by the Wellman-Lord/Allied process and the Cat-Ox process respectively.

Use of SRC (sized to produce fuel for a 500 MW power plant) or low Btu gas is not competitive with stack gas scrubbing.

The use of a large (2000 MW) SRC system improves economics for the process considerably. Fuel costs are about 37¢/MM Btu lower when the size is increased by a factor of four. However, costs are still somewhat higher than flue gas desulfurization costs. A substantial increase in by-product credit (cresylic acid) would be necessary to enable this process to be competitive with stack gas scrubbing.

As the load factor decreases, the unit operating cost for all processes increases. The increase in cost is much greater for SRC and low Btu gas than for flue gas desulfurization thereby reinforcing the previous conclusions.

Total Production Costs - ¢/MM Btu*

	Flue Gas Desulfurization			Solvent Refined Coal		Low Btu Gas
Load Factor	Wet Limestone	Wellman-Lord/Allied	Cat-Ox	500/500	500/2000	
60%	78	83	88	145	108	143
45%	87	95	105	177	128	188

* Coal at $10/T

2.2 1000 MW New Plant (80% Load Factor)

Total Production Costs - ¢/MM Btu

	Flue Gas Desulfurization			Solvent Refined Coal		Low Btu Gas
Coal Cost: $/T	Wet Limestone	Wellman-Lord /Allied	Cat-Ox	1000/1000	1000/4000	
5	44	49	52	80	53	74
10	65	70	73	109	82	108
15	86	90	93	137	110	143

Flue gas desulfurization remains somewhat superior to the use of SRC (sized for 1000 MW) or coal gasification.

The use of a large SRC plant (4000 MW system) improves economics for the process considerably. Costs are about 27¢/MM Btu less than for a 1000 MW SRC plant.

When coal costs are low (about $5/T), the 4000 MW SRC process is competitive with flue gas desulfurization. With coal at $10/T, the 4000 MW SRC process appears to be somewhat more costly than flue gas desulfurization. However, an increase in by-product credit may be realized which could make SRC competitive with flue gas desulfurization.

At a coal price of $5/T, coal gasification costs about 20¢/MM Btu more than the 4000 MW SRC process. As coal prices increase, the spread becomes greater due to lower efficiency of the gasification process. The increase in cost is about 1.3¢/MM Btu for each $ 1/T of coal price increase.

Integration of coal gasification into a combined cycle power plant (utilizing gas turbines and steam turbines) would appear to be more desirable than its use as feed preparation for a conventional power plant.

As the load factor decreases, the unit operating cost for all processes increases as shown by the following table:

Costs - ¢/MM Btu*

	Flue Gas Desulfurization			Solvent Refined Coal		Low Btu Gas
Load Factor	Wet Limestone	Wellman-Lord/Allied	Cat-Ox	1000/1000	1000/4000	
80%	65	70	73	109	82	108
60%	70	76	83	128	92	120

* Coal at $10/T

3. BASES FOR COMPARISON OF DESULFURIZATION TECHNOLOGIES

In this study, an attempt has been made to compare different desulfurization technologies and processes as applied to conventional steam-electric power plants. The processes included in the study are:

1) the wet limestone process
2) the Wellman-Lord/Allied Chemical Process
3) the Cat-Ox Process

flue gas desulfurization

4) solvent refined coal
5) Lurgi coal gasification to produce low Btu gas

Conventional H_2S removal systems are included in the gasification and solvent refined coal processes, a Benfield system for the former and an amine system for the latter. Sulfur is recovered in 2-stage Claus plants. No Claus tail gas treatment facilities are included. An overall sulfur recovery of 90% has been used for flue gas desulfurization processes and low Btu gas, giving an overall SO_2 emission rate of about 0.6 lbs. SO_2/MM Btu. For the solvent refined coal process, the product can typically be desulfurized to about 1.0% sulfur. This produces an SO_2 emission rate of about 1.2 lbs. SO_2/MM Btu from the power plant, but the overall emissions from the solvent refined coal plant plus the power plant would be slightly higher.

Process descriptions and the bases for process designs are given in subsequent sections of this report.

In order to make quantitative comparisons between processes, basic power plant parameters have been established to define the reference plants. These parameters are shown in Table 3.1. Since all control processes depend on the use of coal as fuel or feed, coal data are also listed in the table.

Table 3.1

POWER PLANT PARAMETERS

Power Plant Size	500 MW	1000 MW
Number of boilers	2	4
Size of each boiler, MW	250	250
Age of plant, years	10	new plant
Heat rate, BTU/KWH*	9,500	8,700
Load factor, %	60	80
Electrostatic precipitator	yes	yes
Electrostatic precipitator efficiency, %	98.7	98.7
Minimum gas temperature at stack, ^{o}F	175	175
Coal (fuel or feed)		
HHV (as rec'd.), BTU/LB	12,000	12,000
% sulfur	3.5	3.5
% ash	12	12
% moisture	5	5

* No boiler de-rating used for case of low Btu gas fired boiler

4. FLUE GAS DESULFURIZATION

4.1 Process Descriptions

4.1.1 Wet Limestone

Generally speaking a wet limestone process can be divided into three areas:

Limestone receiving and preparation
Particulate and sulfur dioxide removal
Sludge treating and disposal

Figure 4.1 is a block flow diagram of the wet limestone system.

Limestone arrives as a coarsely ground material, and is conveyed by belt to a storage pile. It then proceeds to a wet ball mill, which produces a limestone slurry that is stored in a slurry feed tank.

The slurry then goes to an SO_2 absorber effluent tank, from which it is circulated to the SO_2 absorber. Overflow from the absorber effluent tank proceeds tc a particulate scrubber. Overflow from this tank is pumped to the sludge disposal pond.

The flue gas first enters a venturi scrubber, where it is sprayed with high velocity limestone slurry. The quenched gas exists from the venturi throat into a sump, where a reduction in gas velocity causes the slurry droplets to fall out.

The particle-free gas then flows upward in the SO_2 absorber, where it contacts countercurrently the limestone slurry. The overall reaction occuring in the absorber is:

$$CaCO_3 + SO_2 \rightarrow CaSO_3 + CO_2$$

Slurry droplets that carry over with the gas are collected on the demister vanes.

The scrubbed gas then enters the reheater, where its temperature is raised to about 175°F. An induced draft fan boosts the pressure of the gas before it enters the stack.

The sludge produced in the process is either pumped directly to a settling pond, or it receives some kind of treatment such as clarification or chemical fixation. Ultimate disposal may be ponding or use as landfill. (20,Pp21-25)

A wet lime process is quite similar to the wet limestone process. Some minor differences do exist, such as the elimination of the grinding step in the lime process. This results in a slightly lower capital investment, and an energy savings. Lime has a higher activity than limestone, so less feed is required, and some energy can be saved in the slurry circulation. On the other hand, lime is several times more costly than limestone, thus offsetting these advantages. Overall, the process designs and costs for lime and limestone systems are very similar.

4.1.2 Wellman-Lord/Allied

The Wellman-Lord/Allied process involves removal of the SO_2 by contacting the flue gas with a sodium sulfite-bisulfite solution. The absorber consists of valve trays, each equipped with a separate scrubbing loop. A prescrubbing section removes fly ash and SO_3. Figure 4.2 illustrates the process.

When the flue gas contacts the sodium sulfite-bisulfite solution, the following reactions occur:

$$SO_2 + Na_2SO_3 + H_2O \rightarrow 2NaHSO_3$$

$$2Na_2SO_3 + SO_3 \rightarrow Na_2SO_4 + Na_2S_2O_5$$

$$Na_2SO_3 + 1/2\ O_2 \rightarrow Na_2SO_4$$

A certain amount of unregenerable salts are formed in the system, and a purge stream is required to control the level of these salts.

Sodium sulfite is regenerated in an evaporator by thermal decomposition of the bisulfite:

$$2NaHSO_3 \rightarrow Na_2SO_3 + SO_2 + H_2O$$

Additional sulfite is generated by reacting make-up sodium hydroxide in this reaction:

$$NaOH + NaHSO_3 \rightarrow Na_2SO_3 + H_2O$$

The sulfite crystals are slurried by the addition of condensate from the wet SO_2 gas purification section following the bisulfite thermal decomposition step.

The compressed SO_2 product gas goes to the reduction area where it is first mixed with natural gas. The preheated mixture flows to a reduction unit which produces S, H_2S, CO_2 and H_2O. After cooling the gas enters a claus unit, where most of the H_2S and remaining SO_2 is converted to elemental sulfur. A coalescer removes the droplets of sulfur, and the tail gas is burned with natural gas, then routed to the scrubber system. (12,pp 131-134)

The conversion of sulfur dioxide to elemental sulfur was chosen for this study because sulfur is usually the most desirable product. Other routes are available, such as conversion to sulfuric acid. This type of plant would perhaps cost less, but the product is not as easy to handle and store as the solid sulfur.

4.1.3 Cat-Ox

Cat-Ox removes SO_2 by oxidation over a vanadium catalyst to SO_3, then absorption of the SO_3 to produce 80% sulfuric acid. The process is illustrated in Figure 4.3.

The use of the catalyst requires a very efficient electrostatic precipitator to prevent clogging. Existing units also require a reheater so that the gas enters the converter at about 890°F. New units should have the converter upstream of the economizer and air preheater, thus avoiding a reheater.

The particle free gas enters the converter, where the SO_2 in the flue gas is oxidized to SO_3. After the economizer and air preheater in a new unit, or cooling water heat exchanger in an existing unit, the SO_3-rich gas contacts a circulating stream of sulfuric acid, which absorbs the SO_3 and water vapor. A very efficient mist eliminator removes entrained acid droplets.

The effluent acid is cooled further, part of it goes to product storage tanks, and part of it returns to the absorption circuit. (12,pp 149-150)

4.1.4 Utility and Energy Consumption

Tables 4.1 and 4.2 compare the utility requirements and energy consumption of the three flue gas desulfurization processes. Figures for the wet limestone and Wellman-Lord/Allied process are from M. W. Kellogg cost models (3, pp 88-91, pp 116-121), while those for the Cat-Ox process are derived from a TVA report (14 p 276).

4.2 Process Complexity

Stack gas scrubbing processes provide relatively simple methods of desulfurization. Table 4.3 lists the major pieces of equipment involved in the three systems.

4.2.1 Wet Limestone and Cat-Ox

In terms of process complexity, wet limestone and Cat-Ox are fairly equal, the following points being considered:

Particulate removal in the wet limestone process is less a concern than in the Cat-Ox process, because the electrostatic precipitator already present in the existing 500 MW plant is adequate for wet limestone particulate control. Cat-Ox requires an additional high-efficiency precipitator to further reduce the particulates level.

SO_2 removal is also somewhat simpler in the wet limestone process, as it is done in one step, rather than the two required in the Cat-Ox process.

Raw material receiving and handling is quite a bit simpler in the Cat-Ox process.

Both processes require special attention given to the mist eliminators. The wet limestone process must provide for adequate washing of the demister to reduce plugging, while Cat-Ox needs a very efficient demister to eliminate a sulfuric acid plume.

Both processes must contend with a product of some sort; wet limestone must dispose of its by-product sludge, while Cat-Ox must store and sell its sulfuric acid product.

4.2.2 Wellman-Lord/Allied

Of the three flue gas desulfurization processes under consideration, Wellman-Lord/Allied is the most complicated.

A rather elaborate circulation system is associated with the SO_2 absorber. Liquid from each tray is removed, then reintroduced at a point above its respective tray.

Generation of the elemental sulfur product is somewhat involved. The steps required to generate the product are evaporation, SO_2 purification, SO_2 reduction, and finally product storage.

In addition to the elemental sulfur product generated, there is a by-product recovered from the purge stream. This consists of sodium sulfate and a small quantity of thionates. Additional process steps are required in the treatment of the purge stream.

4.3 Flexibility of Processes

Application of SO_2 emissions control to the utility industry presents some potential problems due to the nature of the industry. The sizes of the power plants are quite varied. Operation is not a steady affair, with periodic shutdowns occuring from time to time. Flue gas desulfurization appears well suited to these conditions.

Stack gas scrubbing technology can be applied to the wide range of capacities that exists in the utility industry. In Japan existing and planned flue gas desulfurization units applied to utility boilers range from 30 to 500 MW. In the United States units due on-stream by the end of 1974 range from 32 to 820 MW.

Application of flue gas desulfurization to large power plants requires the use of multiple scrubbing trains. For example, the power plants considered in this study use four identical scrubbing trains on the 500 MW plant, and eight identical trains on the 1000 MW plant. The concept of identical trains is most evident in the Cat-Ox process, where $/KW total capital requirement is only about 10 percent higher in a 500 MW unit than a 1000 MW unit. In the wet limestone and Wellman-Allied, doubling the plant size results in about a 30 percent reduction in $/KW total capital requirement. (See section 7.4.1)

Flue gas desulfurization operation follows that of the boiler: when the boiler is down, so is the desulfurization unit. This presents no problems in operating these units, as simple start-up and shut-down accompany the relative process simplicity of flue gas desulfurization. The danger of contamination of the Cat-Ox catalyst by fly ash exists during start-up, so the units are equipped with a start-up by-pass duct, allowing the operation to stablize before actual desulfurization begins. Routine maintenance and cleaning of the systems can be scheduled during boiler down time.

4.4 Status of Technology

Flue gas desulfurization is a commercially proven method of controlling SO_2 emissions. The successful commercial applications of flue gas desulfurization can be separated into eight basic classes (23, p80):

Lime or limestone slurry scrubbing

Sodium sulfite scrubbing with thermal regeneration

Dual media system using dilute sulfuric acid for scrubbing

Double alkali systems

Magnesium oxide

Copper oxide acceptor

Activated Carbon

Once through soda ash solution system

4.4.1 Applications in Japan

Most of the successful commercial installations are located in Japan, where units have been performing adequately for two years or longer, with availabilities of over 95 percent. SO_2 removal efficiencies up to 95 percent have been achieved. Inspired by these successes the Japanese industry has moved ahead vigorously with flue gas desulfurization. The technology has been successfully applied to boiler stack gas, Claus sulfur-plant tail gas, sulfuric acid-plant tail gas, copper smelting tail gas, and iron ore sintering tail gas. (23, pp79-80)

Tables 4.4 and 4.5, and Figure 4.4 show the progress of flue gas desulfurization in Japan. These points are worthy of note (19, pp4,7,10):

There are seventeen wet limestone scrubbing facilities with capacities greater than 20 MW operating in Japan, or scheduled for completion by 1974. Many of the plants attain SO_2 removal levels of 90 percent. The Mitsubishi-Jecco process is commonly used for oil-fired boilers, iron-ore sintering plants, etc., while the Chemico Mitsui and Mitsui Miike processes are used for coal-fired boilers. In addition fourteen lime or limestone processes are to be installed during 1975 and 1976.

Twenty-two double-alkali facilities in units of 20 MW or larger are now operating in Japan or scheduled for completion by the end of 1974, with six other units due onstream by the end of 1975. Seven of the processes use wet absorbents: Nippon Kokan, ammonium sulfite; Chiyoda, sulfuric acid; Kurechi-Kawasaki, Showa Denko, Showa Denko-Ebara, and Tsukishima, sodium sulfite; Kurabo Engineering, sodium sulfate; and Dowa mining, aluminum sulfate. The Hitachi-Tokyo Electric process uses a dry absorbent, activated carbon.

The Wellman-Lord process is currently being used in twelve Japanese locations with sizes of 20 MW or larger. Two additional units are scheduled for completion in 1975. Applications of the process are to industrial boilers, utility boilers, and a Claus furnace.

Three units using magnesium oxide scrubbing are to be in operation by the end of 1974, treating a copper smelter, a sulfuric acid plant, and a Claus unit.

The Sumitomo Shipbuilding process uses dry activated carbon to absorb SO_2 from a utility boiler stack.

Shell's copper oxide process is used on a utility boiler stack gas.

The Mitsubishi - IFP process uses ammonia scrubbing to control SO_2 emissions from two Claus furnaces.

4.4.2 Application in the U.S.

In the United States utility industries, twenty flue gas desulfurization units are expected to be operating by the end of 1974, representing a total of 3481 MW. Many additional units are planned to come onstream before 1980. Tables 4.6 and 4.7 and Figure 4.5 illustrate the planned progress of flue gas desulfurization in the U.S. utilities. The following points are noteworthy: (20,pp49-73)

Fourteen facilities employ lime or limestone scrubbing, the sizes ranging from 30 to 820 MW. Another twenty-eight units are planned for completion by 1980.

Two sodium carbonate scrubbing units are installed at present, with two more units anticipated by 1980.

One magnesium oxide unit is presently installed, and another is planned for completion before 1980.

One double alkali system was started up in March, 1974.

One Cat-Ox unit is currently installed.

Three Wellman-Lord/Allied units are anticipated before 1980.

4.4.3 Operational Problems in the U.S.

Since the introduction of flue gas desulfurization units in the U.S., many problems have plagued their operation and reduced the availability of the units. Most of the chemical problems have been overcome; however some mechanical difficulties still exist. Table 4.8summarizes the operating status of desulfurization units in the United States. (20, pp8-10)

4.4.3.1 Lawrence Power Station

The Lawrence Power Station of Kansas Power and Light has flue gas desulfurization units on its oldest unit, No. 4, and on a unit put into service in 1971, No. 5. No. 4 has a capacity of 115 MW burning natural gas and coal, while No. 5 is rated at 400 MW burning the two fuels.

Both units were built by Combustion Engineering, who also designed and installed the scrubbing system: limestone injection followed by wet scrubbing.

The coal presently burned has a heat content of 12,000 Btu/lb, ash content of 12 percent, and sulfur content of 3.75 percent. Because of the curtailment of strip mining at the Kansas Coal supply site, the feed is to be switched to Wyoming coal, with a heating value of 10,000 Btu/lb, ash content of 10 percent, and sulfur content of 0.4 to 0.8 percent.

When the desulfurization unit on No. 4 began operating in 1968, many problems arose due to improper chemical control of the process. The problems included:

Scale buildup in hot gas inlet ducts

Erosion of scrubber walls and corrosion of the scrubber internals.

Scaling of drain lines, tanks, pumps, marble bed, demister, and reheater.

Scale accumulation on the I.D. fans

Inadequate SO_2 removal due to overburning of limestone in the furnace, and dropout of lime in the scrubber.

After a few months of operation, design had to be modified in these ways:

Installation of soot blowers at the gas inlet duct and reheater.

Raising of the demister.

Running overflow liquor from the pots to the pond.

Installation of a large tank and pump to recirculate the underflow.

These modifications reduced some of the problems with scaling, as well as improving SO_2 removal efficiency.

Other revisions were made in 1970 to further combat the scaling problems, yet demister problem continued, requiring manual washing every other night. In 1972, both modules were modified to use a high solid slurry crystallization process to control saturation and precipitation in the scrubbers.

Since the fall of 1973 performance of the units has improved somewhat. In July and August of 1974 availabilities of near 100% have been reported. Problems still do exist in both modules, with the one on No. 5 experiencing difficulty with poor gas distribution.

Plans are currently underway to convert the unit on No. 5 to tail end wet lime or limestone scrubbing only. Plans for No. 4 include installation of an electrostatic precipitator, and replacement of the scrubbing system. (20, pp16-20)

4.4.3.2 Will County Power Station

The Will County Power Station of Commonwealth Edison has one 167 MW boiler fitted with a flue gas desulfurization unit. The coal fuel has a heating value of 9463 Btu/lb, ash content of 10%, and sulfur content of 2.1%.

In February, 1972, the boiler was fitted with a wet limestone scrubbing system consisting of two modules, referred to as A and B. Table 4.9 summarizes the monthly availabilities of the two modules. (20,p28). Problems with the system soon became apparent. Low wash water pressure contributed to the constant problem of demister plugging. This problem caused both modules to be out of service several days of each month of operation initially. Some modifications in the wash water system yielded no improvements, so the demister elements had to be hand washed, which introduced broken elements into the slurry system.

Module B had to be taken out of service to correct excessive vibration in the reheater section. Both modules experienced additional problems of erosion and plugging of spray nozzles, deposit buildup on venturi nozzles, corrosion cracking, sulfite binding, and fan vibrations.

In 1973, problems with the demister continued, highlighted by the loosening of the demister in Module B, and the subsequent plugging of the reheater by chloride pitting corrosion. In April Module B was taken off stream indefinitely until Module A is satisfactory. A system of continuous underspray and intermittent overspray was installed on Module A to reduce demister plugging.

In 1974 operation of Module A has improved somewhat, although problems still remain, such as: freezing of the venturi throat drive, tank screen blinding, dust corrosion, and vibrations. Some parts of Module B have been used in Module A modifications. (20, pp26-30).

4.4.3.3 Hawthorn Power Station

The Hawthorn Power Station of Kansas City Power and Light uses flue gas desulfurization on Units 3 and 4. Each boiler is rated at 140 MW for natural gas, and 100 MW for coal. Two types of coal are burned: one with heating value of 11,400 Btu/lb, 14% fly ash, and 3% sulfur; the other with heating value of 9800 Btu/lb, 11% ash, and .6% sulfur.

Initially both units employed limestone injection followed by wet limestone scrubbing of the tail gas. After developing plugging in the tubes of Boiler 4 due to limestone injection, the ground limestone was injected into the flue gas near its entry into the scrubber.

Problems encountered by the four identical modules, two per unit, have been similar. Many of the problems have been reduced by process and equipment modifications.

The reaction tank of each module initially had problems with buildup of hard mud in the corners of the tanks. Installation of welded triangle plates and make-up water nozzles near the plates improved the situation.

Plugging problems in the marble bed of the absorber have been solved by installation of stainless steel drain pots with expanded metal covers. The liquid to gas ratio in the scrubber has been increased.

Early spray nozzles used in the units lasted for a very short time, and their frequent replacement was a rather expensive operation. A much cheaper shop-made nozzle also did not last very long. Performance of the nozzles was improved greatly by the use of ceramic nozzles.

The typical demister problems encountered elsewhere have been minimized by the addition of retractable water lance blower under the demister, and by moving the rotary water lance blowers to between the demister vanes.

Availability of the unit on Boiler 3 has increased to about 70 percent, while No. 4 has lagged behind somewhat due to limestone injection related problems. (20, pp36-39)

4.4.3.4 Reid Gardner Power Station

The Reid Gardner Power Station of the Nevada Power Company uses flue gas desulfurization on its two units, each rated at 126 MW. The two sodium carbonate based desulfurization units have operated since March, 1974. Each boiler has a single module unit.

Both units have operated satisfactorily since their start-ups. The operations have been subjected to frequent interruptions due to a lack of the sodium carbonate source, trona. Availabilities during adequate supplies of trona have been 100%, and each unit has operated for 900 hours. None of the problems of scaling, demister plugging, erosion, and corrosion associated with previously mentioned units have surfaced. (20, pp44-45)

4.4.3.5 Cholla Power Station

The Cholla Power Station of the Arizona Public Service Company has a wet limestone scrubbing system installed on its single 115 MW unit. The double-train unit has operated satisfactorily since its start-up in late 1973; however some problems have been evident:

Initial heavy vibrations in the reheat section have been reduced significantly by the installation of baffles to evenly distribute the desulfurized flue gas.

Improper operation of the by-pass damper continues to reduce the scrubber efficiency by allowing some flue gas to pass freely to the stack.

Early problems experienced with the flooded disk, which maintains an equal pressure drop in the system at all loads, have been corrected by eliminating buildup problems around the disk shaft, and by adjustment of the controls.

Some corrosion and plugging problems have been experienced, but they are relatively minor in scale. It is hoped that proper maintenance will keep such trouble spots to a minimum.

In one common trouble area, demister plugging, a new wash water system appears to keep buildup at a minimum. (18, pp3-7)

4.4.3.6 Olin Corporation Sulfuric Acid Plant

Since the startup of this Wellman-Lord unit on the tail gas from a 750 ton per day sulfuric acid plant at Paulsboro, New Jersey, operation has been about what was expected, except for the sodium hydroxide makeup rate, which was initially about 50 percent above design: 3.75 TPD, vs the design rate 2.5 TPD.

Some corrosion problems were present in the early operations, but these problems have been corrected. (24, p43)

4.4.3.7 Wood River Power Plant

The Wood River Plant of Illinois Power Company has a Cat-Ox system installed on its 110 MW unit. The unit has not operated very much since its startup in September, 1972, because of the conversion of the reheat section from natural gas fuel to fuel oil, and problems associated with this conversion. Units on new power plants will not require reheat sections.

During a 24 hour test run, some aspects of the unit operation were noted:

Conversion of SO_2 to SO_3 reached 93%, which is over the guaranteed 90%.

92% removal of SO_2 was achieved, better than the guaranteed 85%.

Acid of 78% concentration was produced when the absorption tower operated at design temperature.

Acid mist leaving the system measured 0.529 mg per cubic foot, compared to the design guarantee of 1 mg/cu ft. (27, pp51-52)

4.4.4 Vendor Capacity

The claim is sometimes made that stack gas scrubbing represents an unfeasible means of meeting the Clean Air Act deadlines, simply because the supplying capability of U. S. manufacturers of scrubbing equipment is inadequate. However, the Industrial Gas Cleaning Institute, which represents over two dozen major U. S. suppliers of SO_2 control technology, has claimed that U. S. suppliers could build 525 systems, averaging 460 MW each within the next seven years. This represents a total of 241,500 MW, which is about 81 percent of the U.S. 1972 total steam-electric generating capacity of about 300,000 MW. This does not consider Japanese suppliers, some of whom are actively seeking U.S. business. (23, p85)

4.5 Environmental Effects

4.5.1 Wet Limestone

4.5.1.1 Sludge Generation

The wet limestone process generates a sludge stream that is composed mainly of $CaSO_3$ and $CaSO_4$. In Japan a further oxidation step yields gypsum ($CaSO_4$) which is marketable; however, in the United States an adequate supply is available, thus the sludge has limited economic use. Table 4.10 shows the average hourly sludge output for four operating wet limestone units, and an approximate composition of the sludge. (17, p6)

Many factors affect the quantity and composition of the sludge, including:

- Size of the power plant
- Type of boiler
- Type of fuel burned
- Sulfur and ash content of the fuel
- Method of fly ash removal
- Method of SO_2 removal
- Stoichiometric ratio of calcium to SO_2
- Efficiency of the SO_2 removal

The calcium compounds in the sludge are mainly calcium sulfite ($CaSO_3$) and calcium sulfate ($CaSO_4$). The relative amount of each depends on the degree of oxidation in the scurbber, which in turn depends on:

- Fly ash content
- pH of the slurry
- Amount of oxygen in the flue gas

4.5.1.2 Disposal Problems

The wet limestone process is known as a throwaway process, because the generated product has no present practical use, and must be disposed of. Disposal causes problems in two areas: large specific volume of the sludge requires a large area for disposal, and dangers of ground water pollution require special attention.

The specific volume of the sludge is a function of its composition. Fly ash alone packs to a volume of about 20 cubic feet per ton of solids, while the sludge packing volume varies from 45 to 75 cubic feet per ton of solid. The former is for sulfates, the latter sulfites. (17, p 8)

A 500 MW plant with an average load factor of 0.6, and burning 3.5% sulfur, would produce a 50% solids sludge at a rate of 1×10^7 cubic feet per year if equipped with a wet limestone scrubbing unit. If the remaining life were 20 years, this unit would generate a total of 2.09×10^8 total cubic feet of sludge. The pond to contain this sludge would be about one-half mile by one-half mile and 37.5 feet deep. (15, p2)

A new 1000 MW plant with a load factor of 80% would produce a 50% solids sludge at the rate of about 2.8×10^7 cubic feet per year, or a total of 8.4×10^8 cubic feet during its 30-year lifetime. This would require a pond of almost one square mile area, and 37.5 feet deep.

If a substantial portion of the U.S. utility industry adopts wet limestone scrubbing, a likewise substantial amount of sludge will be produced. Using the 1972 U.S. total installed steam-electric capacity, and assuming that 54% are coal-fired, yields about 162 million KW of coal-fired capacity. If 50 percent of this capacity, or 81 million KW installs wet limestone scrubbing, and if these plants operate at 60% load factor, then 3.25×10^{10} cubic feet of sludge are generated during 20 years of operation. This is equivalent to about 23 square miles of 50 feet deep sludge. (28, p 53)

Not only the sheer volume of the sludge presents problems in its disposal, but the danger of water pollution also concerns the pollution abatement interests. The presence of $CaSO_3.1/2\ H_2O$ and some trace metals in the sludge, and their availability for leaching by rainwater, poses a potential hazard via ground water pollution. This is not peculiar to sludge disposal however, as ash disposal poses the same type problems. (17, p8)

4.5.1.3 Ultimate Disposal Techniques

It is evident that the one big problem remaining in the technology of wet limestone scrubbing is sludge disposal. Quite an effort is being directed towards rendering the sludge easier to handle by reducing its size and altering its physical properties. Table 4.11 shows the current sludge disposal techniques employed by utilities. (17, p10)

One approach to ultimate disposal of the sludge is chemical fixation, followed by use of the resulting material for landfill. The companies with methods of chemical fixation are: Dravo Corporation, IU Conversion Systems Inc., and Chemfix. (21, P 22-25)

The Chemfix process reportedly can handle a wide range of solid content sludges to produce a soil-like substance, which does not prevent rain water percolation, yet is stable and controls pollution by chemically binding the constituents. The sludge is reacted with sodium silicate and one or more of these settling agents: portland cement, lime, calcium sulfate, and calcium chloride. The result is a gelatine-like material whose hardening time is fixed depending on the pumping time required. The product is acceptable for disposal with no further treating.

The Dravo process treats a wide range of solid content sludges to produce a clay-like material. The sludge is reacted with an admixture called Calcilox just before pumping to the disposal site. Final settling and curing requires about 30 days. The resulting material has been accepted for disposal without containment in at least one site in Pennsylvania.

The IUCS process uses fly ash and lime addition to fix the sludge. In some cases dewatering of the sludge is required before addition of the agents. Testing has shown that the resulting material can develop high strength very quickly. It has also been shown that the combination of trace elements into new crystaline phase can reduce the availability of toxic materials to ground water. The process may also be used to make synthetic aggregate suitable for road base materials.

Another approach to the bulk size reduction and improvement of the physical properties of the sludge is by dewatering, followed by use as landfill. The problem of leachability remains, and covering with clay may be all that is needed to prevent contamination of the ground water.

A sulfate sludge dewatered by filtration or centrifugation may require only additional solar drying to reduce the water content to a level suitable for compaction. A sulfite sludge may require more than simple centrifugation of filtration, such as the addition of dry fly ash to the sludge, or the use of thermal drying. Another solution might be oxidation of the sulfite to achieve the easier dewatering properties associated with sulfate sludges.

The simplest way to dispose of the sludge is by ponding. The sludge is pumped or hauled to a pond designed to contain the raw sludge for a long period of time. The pond is lined with impervious material such as butyl rubber or clay. Leakage detection systems are employed.

4.5.2 Wellman-Lord/Allied and Cat-Ox

4.5.2.1 By-Products

The main product of the Wellman-Lord/Allied desulfurization process is elemental sulfur having a quality suitable for sale to manufacturers of sulfuric acid via the contact process. A 1000 MW unit fired with 3.5 percent sulfur coal, and having a load factor of 0.8 produces about 63,210 tons per year of elemental sulfur. It produces a total of 1.9 million tons in its 30 year life. A 500 MW plant with a load factor of 0.6 produces 20,706 tons per year of sulfur, or about 520,000 tons in the 20 years of remaining life. (12, Appendix B)

A purge stream is required to control the buildup of non-regenerable sodium salts in the process. This stream is concentrated and dried, resulting in an 85 percent Na_2SO_4 product. Davy Powergas is currently researching means of regenerating the sodium ion, so that the amount of purge can be significantly reduced. (13, p5)

The product of the Cat-Ox system is 80% sulfuric acid. A 1000 MW plant burning 3.5 percent sulfur coal, and having a load factor of 0.8, produces about 212,400 tons per year, or 6.4 million tons in a 30 year lifetime, based on 100 percent sulfuric acid. A 500 MW plant with 0.6 load factor generates about 69,600 tons per year 100 percent acid, or about 1.76 million tons in a 20 year remaining life. (25, p46)

4.5.2.2 Marketing Problems

Wellman/Allied and Cat-Ox both produce sulfuric acid related products, 80 percent acid in the Cat-Ox system, and sulfur available for sulfuric acid manufacturing from the Wellman/Allied process.

If a substantial section of the utility industry turns to sulfuric acid producing methods of SO_2 control, the acid production could be quite substantial. For example, sulfuric acid production from desulfurization systems could equal about 60 million tons per year of acid, almost double the U.S. 1972 production of 31.3 million tons. The expected variation in the approach that the utility industry will take to solve the SO_2 pollution problem will dampen the effects on market systems. However, the potential for upset is present. (15, p4)

As a product, elemental sulfur has certain advantages over the 80% sulfuric acid produced by the Cat-Ox process. It is an inert material, thus if necessary, could be dumped with no fear of polluting effects. The specific volume of the sulfur is lower than the acid specific volume, and can be more easily stored.

4.6 Installation

4.6.1 Installation Time

Installation time is of course a function of a varied set of parameters associated with a particular project. In the case of flue gas desulfurization units, the situation facing contractors is the installation of a relatively new technology, sometimes on premises that were constructed with no intent of future additions. These factors tend to extend installation time. As more and more of the units are installed, the acquired experience in this area will probably reduce somewhat the required installation time.

4.6.1.1 Will County Power Station

In September 1970, Babcock and Wilcox was authorized to begin detail engineering of a wet limestone stack gas scrubbing unit for the Will County Power Station of Commonwealth Edison. Completion of the project was set at December 31, 1971, by the Illinois Commerce Commission. It was apparent that orders for major equipment items would have to be placed early to meet the deadline, so authorization to purchase was given on September 28, 1970. By July, 1971, most of the major equipment items were at the site.

Erection was scheduled to begin on April 1, 1971, but it was delayed for six weeks because it was discovered that a slab type of foundation would not support the scrubber system.

In spite of the late start, the unit was essentially complete by the end of February, 1972, one year and seven months after the beginning of detail engineering. (26, p92)

4.6.1.2 Phillips Power Station

In the case of the Phillips Power Station of the Duquesne Light Company, the contract was awarded to the Chemico Corporation in July, 1971, and Duquesne felt that a reasonable target date for completion would be July, 1973, which was 34 months after the decision to install a scrubbing system. Under pressure from the State of Pennsylvania, the target date for completion was set at January 1, 1973.

On July 9,1973, six months late, the first portion of the unit was completed. Several reasons for the delay were: some development engineering was required to adapt the scrubbers, some suppliers were late with their orders, and some major delays occurred in the field. (22, pp3-5)

4.6.1.3 Wood River Power Plant

The original schedule for the Cat-Ox unit at the Wood River plant of Illinois Power Company called for design and cost estimates to begin in June, 1970, with detail engineering and procurement to be initiated in November, 1970. The electrostatic precipitator unit was to be placed in operation early in 1972, and the complete flue gas desulfurization unit operational early in 1974.

Some construction delays occurred, then the natural gas shortage prompted a redesign of the reheat section to operate on fuel oil. A change to external burners was also found to be necessary.

The reheat section will not be needed on a unit installed on a new plant. (27, pp 51-52)

4.6.1.4 A Wellman-Lord/Allied Unit

Davy-Powergas has been awarded a contract to construct a Wellman-Lord/Allied desulfurization system on a new power plant. The project was awarded in the Spring of 1974, and engineering, procurement, and construction is expected to take thirty to thirty-three months. (13, p4)

4.6.2 Space Requirements

One of the major factors to be considered in anticipating a flue gas desulfurization system is the amount of space the unit will require. This factor is especially important when considering retrofitting a system in a rather congested power plant.

4.6.2.1 Wet Limestone

In a wet limestone unit the main space-taking elements are the limestone pile and related equipment, scrubbing trains, and the sludge disposal facilities. The requirements of these three areas for a 500 MW unit are shown below: (22, pp88-89)

Limestone pile and slurry preparation	76,800 ft^2
Four scrubbing trains (side by side)	30,888 ft^2
Sludge disposal pond	153,000,000 ft^2

The sludge pond is fifty feet deep.

Sometimes there is simply not enough space at the power plant for on-site disposal of the sludge. In this case a thickener may be used, along with a small pond. The thickened sludge is then hauled away for disposal.

A new 1000 MW unit requires 8 scrubbing trains to process the stack gas, twice the limestone pile that a 500 MW unit needs, and 2.67 times the sludge pond area. The space requirements are shown below:

Limestone pile and slurry preparation	153,720 ft^2
8 scrubbing trains	61,776 ft^2
Sludge pond	409,000,000 ft^2

4.6.2.2 Wellman-Lord/Allied

In the Wellman-Allied process, the main areas are the flue gas scrubbing trains, and the SO_2 reduction-regeneration and purge treatment area. Space requirements for a 500 MW unit are: (12, pp137-139)

4 scrubbing trains	34,690 ft^2
Reduction-regeneration and purge	43,420 ft^2

A 1000 MW plant requires twice the space for the scrubbing trains that a 500 MW unit does; however, the reduction-regeneration area will not increase its size linearly with capacity because larger tanks and vessels can be used. Assuming a size increase in this area governed by a 0.5 exponent yields the following space requirements for a 1000 MW unit:

8 scrubbing trains	69,380 ft^2
Reduction, regeneration, purge	61,400 ft^2

Not included in the space requirements for the Wellman-Lord/Allied system is a required sulfur product storage area. This area, including tanks and a dike surrounding them, will likely require an area equal to about one-sixth the process area.

4.6.2.3 Cat-Ox

The cat-Ox system has its space requirements in two main areas: the actual process area, and storage space for the produced sulfuric acid. For a 500 MW plant the areas are: (25, p46)

Process	47,340 ft^2
Acid Storage	52,600 ft^2

A new 1000 MW plant requires:

Process	115,550 ft^2
Storage	96,300 ft^2

4.6.3 Retrofitting Problems

Some power plants have adequate open areas near the stack that can be used for scrubbing trains or converters, and areas available elsewhere to locate other facilities. In these cases retrofitting is not a large problem, and the installation costs associated with these units is not much more than a comparable installation in a new plant.

The arrangement of some power plants is such that retrofitting a flue gas desulfurization unit presents a major problem. Retrofitting at these sites is likely to require large expenditures for extra ductwork, foundations, steel structures, and the relocation of existing equipment, buildings, railroad tracks, etc.

As examples:

Installation of a wet limestone scrubbing system at the W.H. Sammis Power Plant of the Ohio Edison Co. would require moving of about 1000 feet of railroad track, relocation of a considerable portion of the coal pile, and the installation of considerable extra ductwork on three of the units. (29, p9)

Retrofitting a wet limestone system at the Eastlake Power Plant of the Cleveland Electric Illuminating Co. would require demolition and reconstruction of two service buildings, relocation of a silo and part of a conveyor, and rerouting of a pipe bridge. (29, p13)

4.6.4 Time Out of Service for Retrofitting

Although the actual construction effort involved in retrofitting a flue gas desulfurization unit may be quite extensive, the actual tie-in time that is required is not exceedingly long.

If planned properly, tie-in of the scrubber duct to the boiler duct can be done in two to three weeks. As this is the amount of time boilers are usually shut down for maintenance, tie-in can usually be accomplished in a way that minimizes power plant outage. (29, p8)

TABLE 4.1

MAJOR EQUIPMENT AREAS FOR 500 MW NEW FGD SYSTEM *

SCRUBBER	FEED PREPARATION	PARTICULATES REMOVAL	SO_2 REMOVAL	GAS HANDLING, REHEAT & HEAT RECOVERY	PRODUCT TREATING	BY-PRODUCT TREATING & REGENERATION
Wet Limestone	1 Wet Ball Mill	4 Venturi Scrubbers 4 Venturi & MBA Sumps	4 SO_2 Scrubbers 4 Effluent Hold Tanks Isoprene Lining 10 Slurry Recycle Pumps 40 Soot Blowers	4 Gas Reheaters 4 ID Fans		
Wellman-Lord/ Allied		4 Venturi Scrubbers	4 SO_2 Scrubbers	4 Gas Reheaters 4 ID Fans	1 SO_2 Reduction Unit	1 Chiller Crystallizer Tank 1 Centrifuge 1 Rotary Dryer 1 Dust Collector 2 Evaporator-Crystallizer
Cat-Ox		4 High T. High Eff. Precipitators	4 SO_2 Convertors 2 Acid Absorbers and mist Eliminators 6 Acid Circulation Pumps	4 Fluid/Air Heaters 4 ID Fans	4 Condensate Heaters 6 Acid Circulation Pumps	

* Total Cost Greater than 100 M 1975 $

TABLE 4.2

ANNUAL UTILITY CONSUMPTION BY FGD PROCESSES

	WET LIMESTONE		WELLMAN-ALLIED		CAT-OX	
	500 MW	1000 MW	500 MW	1000 MW	500 MW	1000 MW
Electricity KWH	73.3×10^6	179×10^6	68.9×10^6	168.4×10^6	58.8×10^6	187×10^6
Steam MLB	-	-	1.612×10^5	3.94×10^6	-	(1.28×10^6)
Fuel Oil GAL	2.36×10^6	5.77×10^6	2.36×10^6	5.77×10^6	8.057×10^6	-
Natural Gas MCF	-	-	433,400	1,059,000	-	-
Process Water MGAL	136,000	332,500	19,000	46,500	-	-
Cooling Water MGAL	-	-	6,160,000	15,045,000	6,687,000	680,000

TABLE 4.3

ANNUAL ENERGY CONSUMPTION BY FGD PROCESSES

	WET LIMESTONE		WELLMAN-ALLIED		CAT-OX	
	500 MW	1000 MW	500 MW	1000 MW	500 MW	1000 MW
Electricity KWH	73.3×10^6	179×10^6	68.9×10^6	168.4×10^6	58.8×10^6	187.0×10^6
Equivalent BTU	6.96×10^{11}	1.56×10^{12}	6.55×10^{11}	1.47×10^{12}	5.59×10^{11}	1.63×10^{12}
Steam MLB	-	-	1.612×10^6	3.94×10^6	-	(1.28×10^6)
Equivalent BTU	-	-	2.1×10^{12}	5.1×10^{12}	-	(1.67×10^{12})
Fuel Oil GAL	2.36×10^6	5.77×10^6	2.36×10^6	5.77×10^6	8.057×10^6	-
Equivalent BTU	3.54×10^{11}	8.66×10^{11}	3.54×10^{11}	8.66×10^{11}	1.2×10^{12}	-
Natural Gas MCF	-	-	433,400	1.059×10^6	-	-
Equivalent BTU	-	-	4.33×10^{11}	1.06×10^{12}	-	-
Total BTU	1.05×10^{12}	2.43×10^{12}	3.54×10^{12}	8.50×10^{12}	1.76×10^{12}	-
Total Heat Rate	10,184	9,047	11,139	10,157	9,717	8,700

Table 4.4

Flue Gas Desulfurization in Japan

(REF. 19)

Process	Application Through 1974	Application From 1975-76	Size Range MW
Wet Lime-Limestone Scrubbing	7 Utility Boilers	9 Utility Boilers	30-500
	4 Industrial Boilers	2 Industrial Boilers	22-175
	1 Copper Smelter		29
	3 Sintering Plants	3 Sintering Plants	26-279
	1 Heating Furnace		32
	1 Diesel Engine		62
Double-Alkali H_2SO_4 Absorbent	1 Utility Boiler	1 Utility Boiler	250-350
	5 Industrial Boilers	1 Industrial Boiler	27-230
Double-Alkali Na_2SO_3 Absorbent	1 Utility Boiler	2 Utility Boilers	150-450
	9 Industrial Boilers	1 Industrial Boiler	25-150
	2 Sulfuric Acid Plants		37-43
Double Alkali $(NH_4)_2SO_3$, $(NH_4)_2SO_4$, $Al_2(SO_4)_3$, Carbon Absorbent	1 Utility Boiler		150
	1 Industrial Boiler	1 Industrial Boiler	31-53
	1 Sintering Plant		46
	1 Sulfuric Acid Plant		82
Wellman-Lord	1 Utility Boiler		250
	10 Industrial Boilers	2 Industrial Boilers	50-400
	1 Claus Furnace		23
Magnesium Oxide	1 Copper Smelter		28
	1 Sulfuric Acid Plant		25
	1 Claus Unit		162
Carbon, Copper Oxide, Ammonia	1 Industrial Boiler		37
	2 Claus Furnaces		3-14

Table 4.5

Total Flue Gas Desulfurization Figures

For Japan

(REF. 19)

Application	Number of Sites Through 1974	Number of Sites 1975-76	Total MW Through 1974	Total MW 1975-76
Utility Boilers	12	12	1816	3734
Industrial Boilers	30	7	2884	663
Sulfuric Acid Plants	3	-	162	-
Claus Units	4	-	202	-
Sintering Plants	4	3	341	534
Others	4	-	151	-
Totals	57	22	5556	4931

Table 4.6

Flue Gas Desulfurization Units On Stream by 1974 On U.S. Utilities

(REF. 20)

Location	Process	Size MW	Type*	Fuel	%S	Year Completed
Arizona Public Service Cholla No. 1	Limestone	115	R	Coal	.44	1973
City of Key West Key West Power Plant	Limestone	37	N	Oil	2.4	1972
Commonwealth Edison Will County No. 1	Limestone	167	R	Coal	.6-3.0	1972
Dairyland Power Co-op. Alma Station	Lime Injection	80	R	Coal	3.0-3.5	1971
Detroit Edison St. Clair No. 6	Limestone	180	R	Coal	3.7	1974
Duquesne Light Phillips	Lime	410	R	Coal	1.0-2.8	1973
General Motors Chevrolet Parma 1,2,3,4	Double Alkali	32	R	Coal	2.5	1974
Illinois Power Wood River No.4	Cat-Ox	110	R	Coal	2.9-3.2	1972
Kansas City Power & Light Hawthorn No. 3	Limestone Injection & Scrubbing	140	R	Coal	.6-3.0	1972
Kansas City Power & Light Hawthorn No. 4	Limestone Injection & Scrubbing	100	R	Coal	.6-3.0	1972
Kansas City Power & Light La Cygne No. 1	Limestone	820	N	Coal	5.2	1973
Kansas Power & Light Lawrence No. 4	Limestone Injection & Scrubbing	125	R	Coal	3.5	1968
Kansas Power & Light Lawrence No. 5	Limestone Injection & Scrubbing	400	N	Coal	3.5	1971
Louisville Gas & Electric Paddy's Run No. 6	Lime	65	R	Coal	3.5-4.0	1973
Nevada Power Reid Gardner No. 1	Sodium Carbonate	125	R	Coal	.5-1.0	1973
Nevada Power Reid Gardner No. 2	Sodium Carbonate	125	R	Coal	.5-1.0	1973
Potomac Electric & Power Dickerson No. 3	Mag-Ox	100	R	Coal	2.0	1973
Southern California Ed. Mojave No. 1	Limestone	160	R	Coal	.5- .8	1974
Southern California Ed. Mojave No. 2	Lime	160	R	Coal	.5- .8	1973
Tennessee Valley Authority Shawnee No. 10	Lime/Limestone	30	R	Coal	-	1972

* N=New
R=Retrofit

Table 4.7

Planned Flue Gas Desulfurization Units on U.S. Utilities (1975-1980)

(REF. 20)

Process	Type	Number	Total MW
Limestone Scrubbing	N	25	8576
Limestone Scrubbing	R	3	1002
Lime Scrubbing	N	7	3950
Lime Scrubbing	R	17	5560
Lime/Limestone	N	8	3748
Lime/Limestone	R	2	830
Sodium Carbonate	N	1	125
Sodium Carbonate	R	1	125
Wellman/Allied	N	1	375
Wellman/Allied	R	2	455
Mag-Ox	N	-	-
Mag-Ox	R	1	120

Table 4.8

Status of U.S. Utilities : Flue Gas Desulfurization Units

(REF. 20)

Utility and Unit	Status
Arizona Public Service Cholla No. 1	Some mechanical problems such as frequent reheat section vibration, unit availability consistently above 90%
Boston Edison Mystic No. 6	System shut down indefinitely due to lack of EPA funding. Recent availabilities are: March - 87%, April - 81%, May - 57%, June - 80%
Commonwealth Edison Will County No. 1	Module A availabilities are: April - 73%, May 93%, June - 54%, July - 95%, August - 91%, September - 85%; Module B is down until Module A is satisfactory
Dairy Power Co-op. Alma Station	Demonstration unit with longest run of two days
Duquesne Light Phillips Station	Only about 40% of stations capacity is treated because fly ash overloads the clarifier. Operating time for Modules 1-4 have been 1756, 762, 815, 1707 hours
General Motors Parma Plant	Availability has been 100% since April, but only two modules have operated at a time because of low demand
Illinois Power Wood River	Unit operated 700 hours in last two years because of conversion of reheater to fuel oil
Kansas Power & Light Hawthorn No. 3	Availability has increased from 30% in 1973 to 70% recently
Kansas Power & Light Hawthorn No. 4	Converted from injection to tail end scrubbing. More problems encountered here than with Unit 3
Kansas Power & Light LaCygne	Many initial deposit problems due to poor P_H control. Recently availability is ≈80% with weekly cleaning of each module
Kansas Power & Light Lawrence No. 4	SO_2 removal is only 75% and daily automatic, weekly manual washing is required. Precipitator and system to be replaced in 1977
Kansas Power & Light Lawrence No. 5	Many of No. 4's problems encountered here as well as poor gas distribution
Louisville Gas & Electric Paddy's Run	Availability near 100%; however, since unit is on a peaking boiler, many runs do not justify start-up
Southern California Edison Mojave No. 2	Availability over 80%
Nevada Power Reid Gardner No. 1	Availability over 90% until Na_2CO_3 supply diminished
Potomac Electric & Power Dickerson No. 3	Prior to August, unit was frequently down, but no record kept. Since August availability is 34%

Table 4.9

Will County No. 1 Flue Gas Desulfurization Availability

(REF. 20)

Period	Availability %		Period	Availability %		Period	Availability %	
	A	B		A	B		A	B
March 1972	0	35.4	January 1973	0	0	January 1974	0	0
April	33.9	13.7	February	21.9	24.3	February	0	0
May	69.5	31.8	March	64.8	10.7	March	20.9	0
June	8.4	30.9	April	6.2	13.1	April	72.3	0
July	0	0	May	0	0	May	93.1	0
August	78.7	20.6	June	.9	0	June	54.5	0
September	0	29.5	July	51.4	0	July	95.8	0
October	0	0	August	19.2	0	August	91.3	0
November	0	0	September	0	0			
December	21.8	29.7	October	32.0	0			
			November	50.8	0			
			December	0	0			

Table 4.10

Characteristics of Sludge from Wet Limestone Units

(REF. 17)

Station	Sludge Output Rate LT/hr (Dry Basis)*	Sludge Composition Dry Basis (wt%)				Estimated Solids Content of Dewatered Sludge Wt.%
		$CaSO_3 \cdot 1/2\ H_2O$	$CaSO_4 \cdot 2H_2O$	$CaCO_3$	Fly Ash	
Will County No. 1	17.5	50	15	20	15	35
Key West Power St.	2.4	20	5	74	1	50
La Cygne	12.5	12.5	40	30	15	35
Cholla	3.1	15	20	-	65	50

*LT = long tons

Table 4.11

Sludge Disposal in U.S. Utilities

Facility	Effluent Management	Clarifier	Pretreatment		Chemical	Ultimate Disposal	
			Vac-Filter	Pond		Pondry	Landfill
Lawrence 4 & 5	Closed Loop	--	--	X	--	X	--
Hawthorn 3 & 4	Open Loop[a]	X	--	X	--	X	--
Will County 1	Closed Loop	X	--	--	X[d]	--	X
Key West	Open Loop	--	--	X	--	X	--
La Cygne	Closed Loop	--	--	X	--	X	--
Cholla	Open Loop[b]	--	--	X	--	X	--
Paddy's Run	Closed Loop	X	X	--	--	--	X
Phillips	Closed Loop	X	--	X	X[e]	X	X
Mojave 2	Closed Loop	X	--	X	X[e]	X	X
Parma	Closed Loop	X	X	--	--	--	X

a) Closed loop for clarifier, open loop for pond

b) Solar evaporation

c) Aided by solar evaporation

d) Chicago fly ash method

e) Dravo fly ash method

FIGURE 4.1 WET LIMESTONE PROCESS FLOWSHEET

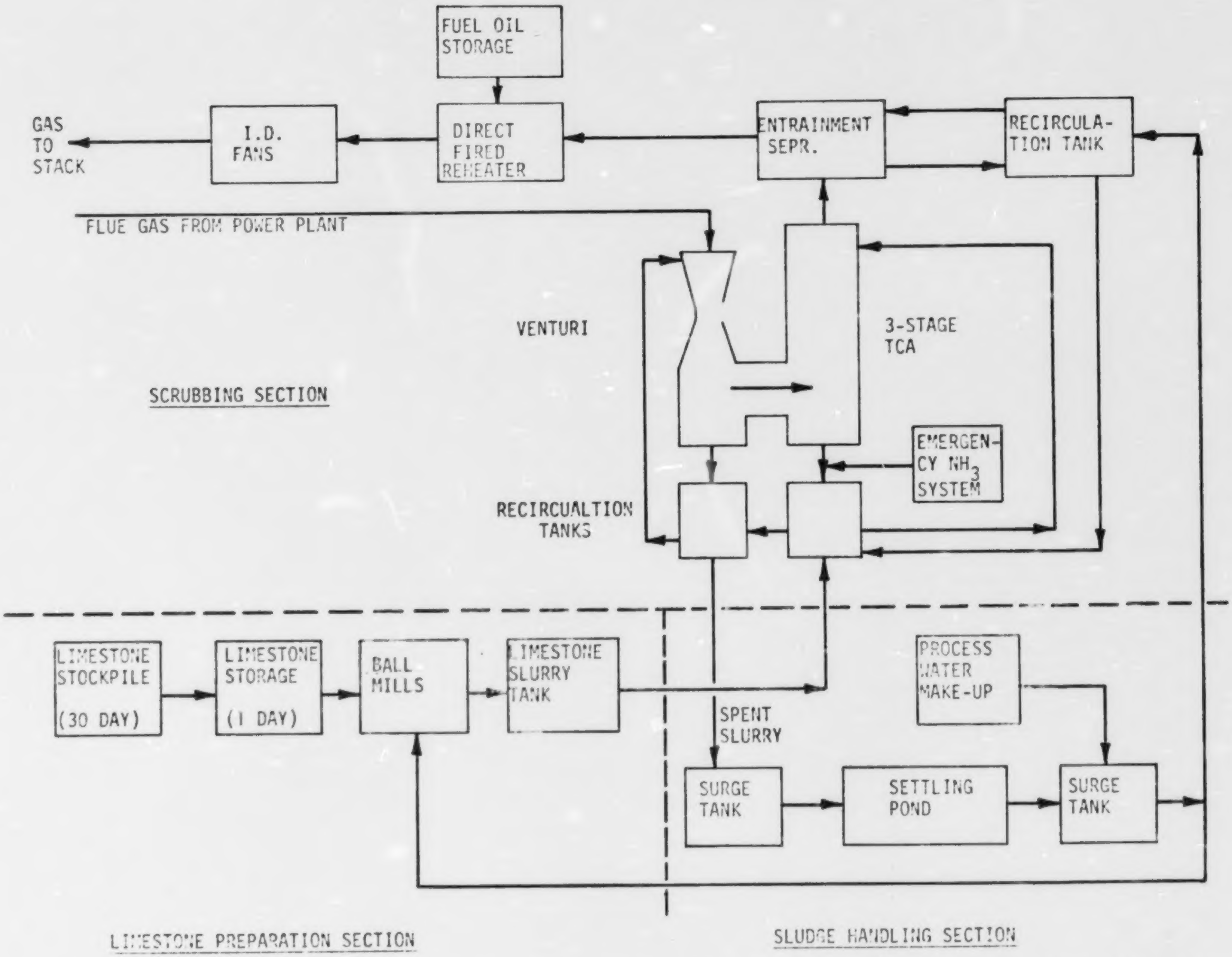

FUEL OIL STORAGE
GAS TO STACK
I.D. FANS
DIRECT FIRED REHEATER
ENTRAINMENT SEPR.
RECIRCULATION TANK
FLUE GAS FROM POWER PLANT
VENTURI
3-STAGE TCA
SCRUBBING SECTION
EMERGENCY NH_3 SYSTEM
RECIRCUALTION TANKS
LIMESTONE STOCKPILE (30 DAY)
LIMESTONE STORAGE (1 DAY)
BALL MILLS
LIMESTONE SLURRY TANK
SPENT SLURRY
PROCESS WATER MAKE-UP
SURGE TANK
SETTLING POND
SURGE TANK
LIMESTONE PREPARATION SECTION
SLUDGE HANDLING SECTION

FIGURE 4.2

WELLMAN-LORD/ALLIED PROCESS FLOWSHEET

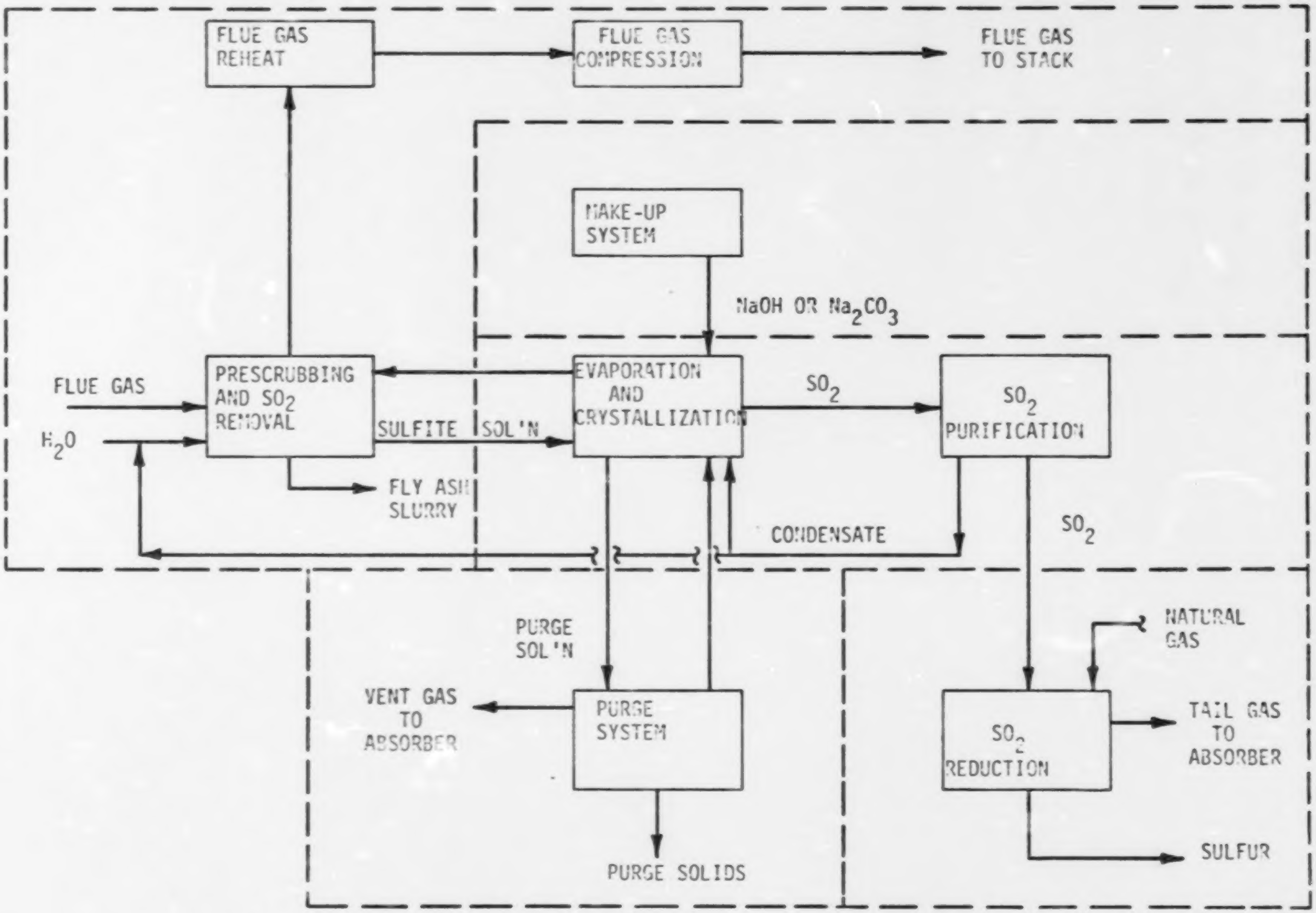
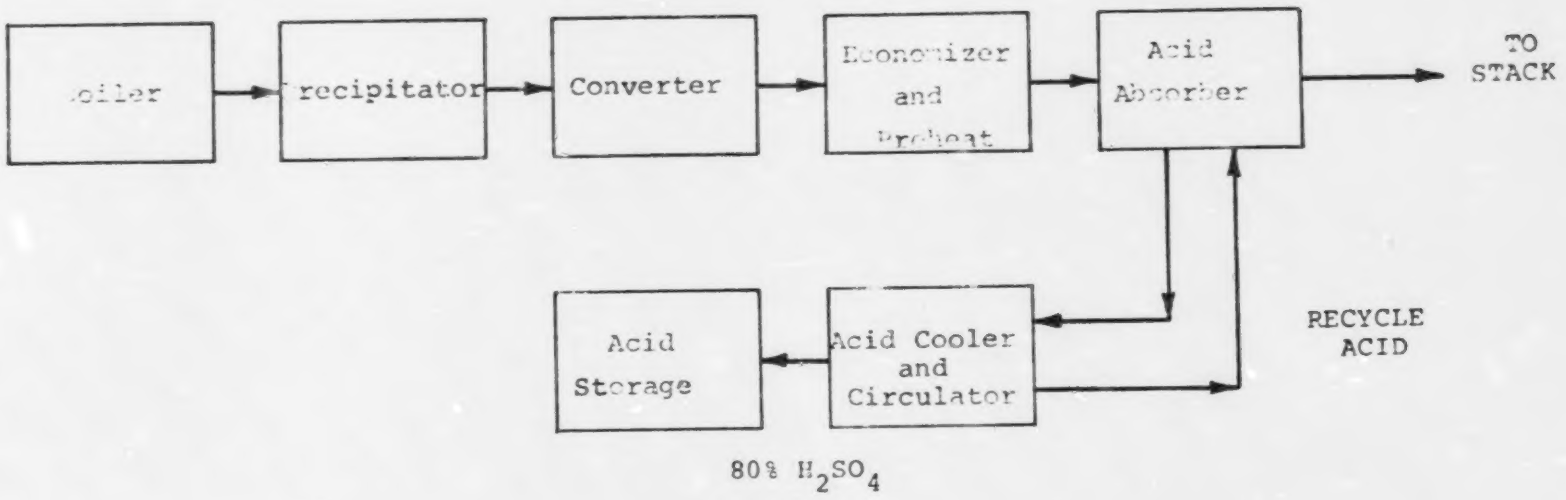

FLUE GAS REHEAT
FLUE GAS COMPRESSION
FLUE GAS TO STACK
MAKE-UP SYSTEM
NaOH OR Na_2CO_3
FLUE GAS
H_2O
PRESCRUBBING AND SO_2 REMOVAL
SULFITE SOL'N
EVAPORATION AND CRYSTALLIZATION
SO_2
SO_2 PURIFICATION
FLY ASH SLURRY
CONDENSATE
SO_2
PURGE SOL'N
NATURAL GAS
VENT GAS TO ABSORBER
PURGE SYSTEM
SO_2 REDUCTION
TAIL GAS TO ABSORBER
PURGE SOLIDS
SULFUR

FIGURE 4.3
CAT-OX PROCESS FLOWSHEET
(INTEGRATED SYSTEM)


Boiler
Precipitator
Converter
Economizer and Preheat
Acid Absorber
TO STACK
Acid Cooler and Circulator
RECYCLE ACID
Acid Storage
80% H_2SO_4

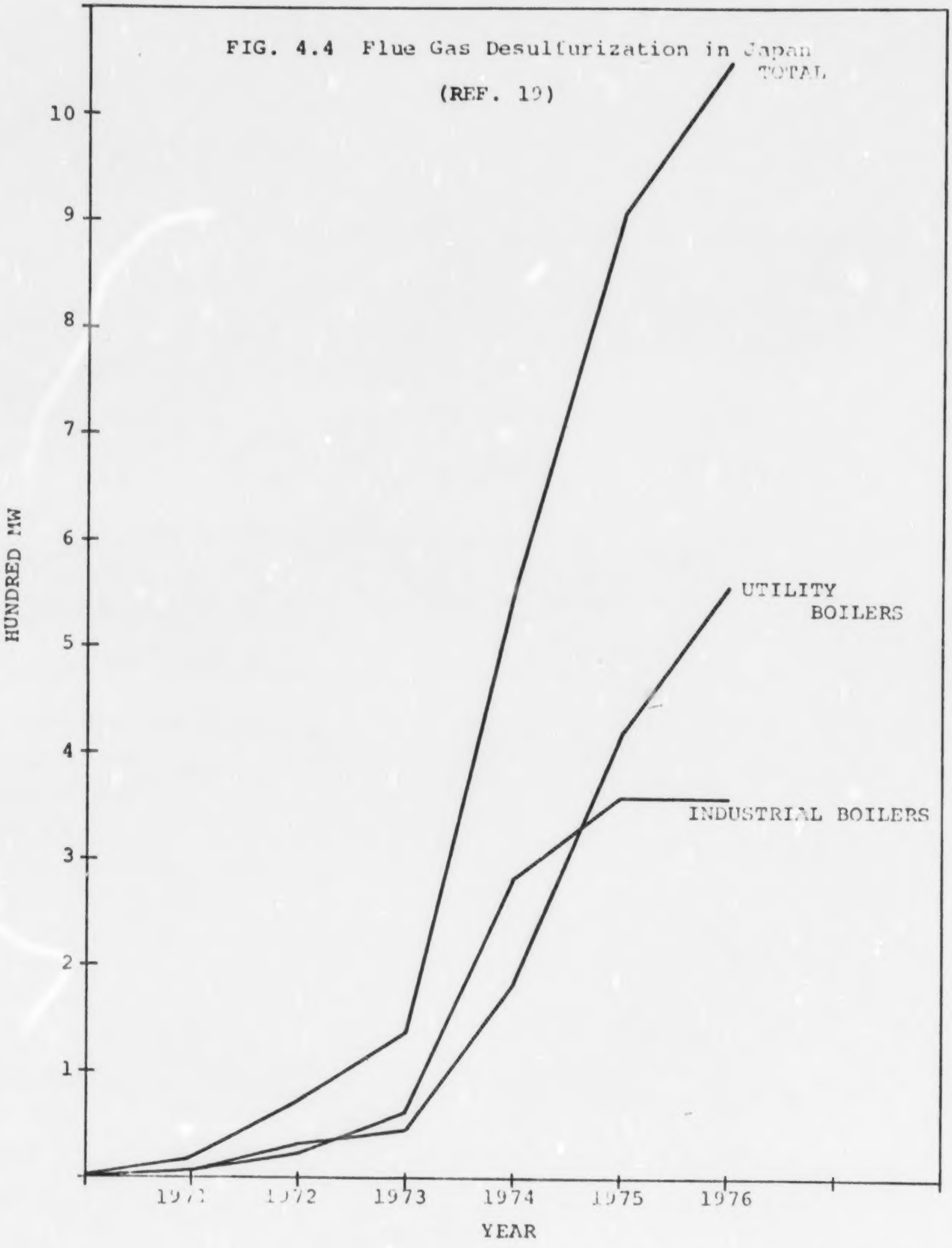


FIG. 4.4 Flue Gas Desulfurization in Japan

(REF. 19)

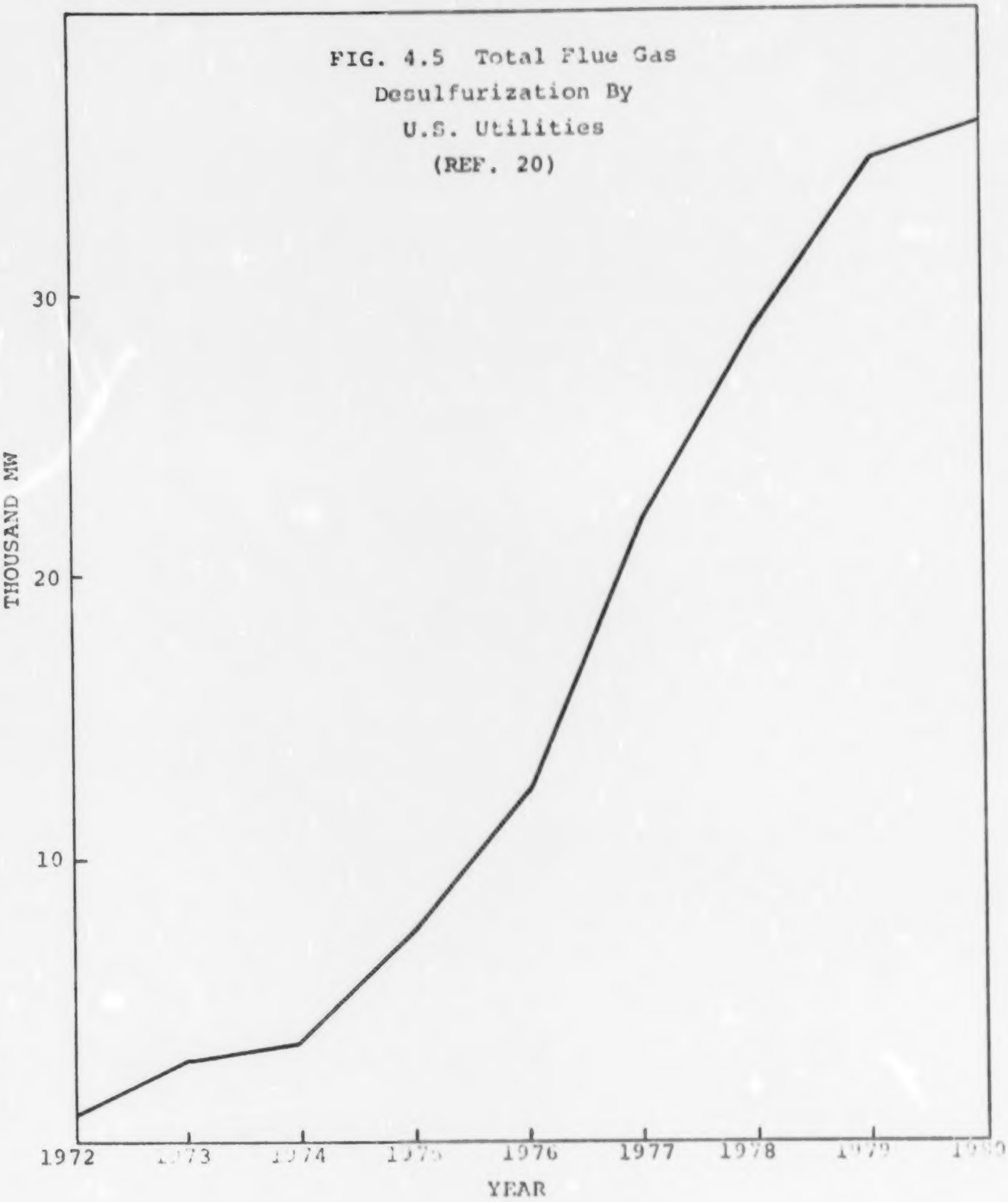


FIG. 4.5 Total Flue Gas Desulfurization By U.S. Utilities (REF. 20)

5. Solvent Refined Coal

5.1 Process Description

The solvent refined coal (SRC) process evaluated in this study is based on the Stearns-Roger Corporation design performed for the Pittsburg and Midway Coal Mining Company (2, Appendix A).* Some modifications to the process have been made (3, pp. 198-199). The plant is designed to operate at steady state conditions for 340 days/year (93.2% onstream factor). Since SRC is easily stored and shipped, SRC plants need not be integrated with a single power plant, but could serve several power plants within a given geographical area. For this study, two different size SRC plants have been considered. The first is sized to produce fuel corresponding to a particular power plant consumption. The second is assumed to be four times this size. It is convenient to identify the size of the SRC plants in terms of the equivalent power production from use of the solvent refined coal. Thus, four different size SRC plants were considered:

500 MW
2000 MW - serving a 500 MW power plant

1000 MW
4000 MW - serving a 1000 MW power plant

Salient consumption/production figures are given in the following table:

Power Plant Size	1000 MW System		500 MW System	
SRC Plant Size	4000 MW	1000 MW	2000 MW	500 MW
SRC Product: 10^9 BTU/D	717.2	179.3	293.7	73.4
Coal Feed: T/D	39,016	9,754	15,972	3,993
Sulfur Product: LT/D	861	215	352	88.1
Cresylic Acid Product: T/D	488	122	200	49.9

---

* Appendix A of reference 2.

The SRC product from the plant can be produced as a solid or a liquid. It is expected to have the following properties:

HHV (BTU/LB)	15,960
% S	0.6 - 1.0
% Ash	0.1 - 0.2

The process has been divided into nine different sections. A brief description of each section follows:

5.1.1 Section 1 - Coal Handling and Grinding

(2, App. A, pp 3-2 to 3-14)

Raw coal from storage is crushed to reduce the coal particle size to < 1/8" (1/8 x 0). Oversize coal is recirculated to the crushers. The fine coal is then processed through flash dryers to remove moisture. Wet coal drops countercurrent to rising hot flue gases from the dissolver preheater waste heat boilers.

5.1.2 Section 2 - Slurry Preheat and Dissolvers

Coal is slurried with solvent (anthracene oil) at the proper ratio (about 2:1 to 3:1 solvent to coal) and pumped through preheat exchangers. Hydrogen is added to the slurry and it is then passed through the dissolver preheaters and dissolvers. The coal dissolves in the anthracene oil in the presence of hydrogen at a pressure of 1000 psig and a temperature of 825°F. The dissolution of coal involves hydrogenation and depolymerization. The coal depolymerization and dissolving process begins in the preheater where the material goes through a gel stage and dissolution is completed to equilibrium in the dissolver (1, p. 17). Heavy oils, hydrocarbons, H_2S and CO_2 are formed. Undissolved material consists of the ash content of the coal.

Effluent from the dissolvers is cooled by heat exchange with slurry feed and then combined into a single stream before entering the high pressure flash vessel where vapor and liquid are separated. The vapor stream containing light hydrocarbons, phenols, cresols, water vapor, CO_2, and fuel gas is fed to the gas-liquid separation portion of the plant (Section 4).

The make-up and recycle hydrogen compressors are included in this section.

5.1.3 Section 3 Ash Filtering and Drying

Slurry from the high pressure flash vessel flows to the filter feed vessel and then to the rotary precoat filters. These units which operate at 150 psig and 600°F are used to separate the ash residue and undissolved carbon from the SRC-solvent solution. The filter cake is washed with light solvent to remove SRC-solvent solution. The ash portion containing some carbon is transferred to the ash drying section for further solvent recovery and then to storage.

5.1.4 Section 4 - Solvent, Light Oil, and Cresylic Acid Recovery

Vapor from the high pressure flash vessel (Section 2) is cooled and partially condensed in a series of heat exchangers. Vapor-liquid separation occurs in the high pressure condensate separator. Gas from the separator is used for power recovery in an expansion turbine before flowing to Section 7. A portion of the gas is blended with make-up hydrogen and sent via the hydrogen recycle compressor to join the slurry at the preheater inlet.

The water phase (containing phenols) from the high pressure condensate separator is sent to the phenol and cresylic acid recovery unit. The organic phase flows to the first stage high pressure condensate flash drum.

The condensate from this vessel is combined with filter vent gas and flashed in the intermediate flash vessel. Remaining liquid is flashed again in the low pressure flash vessel. Gas streams from the two flash vessels (rich in H_2, CO_2, H_2S, and C_1 - C_3) are compressed and sent to Section 7.

The liquid stream from the low pressure flash vessel is preheated and sent to the fractionation area. This area consists of two fractionation towers--the wash solvent splitter column and the light ends column. In the wash solvent splitter column, anthracene solvent is removed from the bottom and recycled to the coal slurrying section. Wash solvent and lighter materials are removed overhead and sent to the light ends column. Here, wash solvent is removed from the bottom and returned to the filters in Section 3. Light hydrocarbons and light oils (containing cresylic acid) are removed overhead and sent to the cresylic acid recovery unit.

Filtrate (containing the SRC) from the filtrate separator is sent to the vacuum preflash vessel for removal of light materials. These compounds are sent to Section 4. The liquid phase is preheated in a fired heater and pumped to the vacuum flash vessel. Vapors leaving the vacuum flash vessel flow through a series of exchangers for heat recovery and then to the fractionation area. Liquid from the vacuum flash vessel is the solvent refined coal. It may be used in this form (the temperature must be maintained above 300°F) or it may be transferred to Section 5 to be solidified.

5.1.5 Section 5 - Product Solidification

This section is necessary if it is desired to produce a solid product. Liquid SRC is transferred to flaking drums and solidified using cooling water. The solid product is transferred to storage silos via conveyors. From here it can be loaded on rail cars or barges for shipment.

5.1.6 Section 6 - Hydrogen Plant

Hydrocarbons and light oil by-product streams are used as hydrogen plant feed. No natural gas is imported. A conventional steam reforming unit followed by shift conversion is used to produce the required amount of hydrogen. Steam produced is used elsewhere in the process and CO_2 produced is vented.

5.1.7 Section 7 - Sulfur Removal and Recovery

The H_2S removal facility is a conventional regenerative amine unit. Feed for the unit consists of the following streams:

- H_2-rich vent gas from the expander
- First stage high pressure condensate flash vapor
- Intermediate flash vessel vapor
- Low pressure flash vessel vapor
- Light ends column overhead vapor

H_2S is absorbed in amine solution in the absorber and stripped out in the stripper. Desulfurized gas from the unit is used as plant fuel. The H_2S rich gas (stripper overhead) is sent to a conventional Claus unit for the production of elemental sulfur.

5.1.8 Section 8 - Steam and Power Generation

The plant is in energy balance with respect to steam and electric power consumption. Steam generation and electric power generation facilities are included. Fuel gas and light fuel oil produced by the process are used in the boilers.

5.1.9 Section 9 - Other Offsites

This section includes other offsites--water treating, the cooling tower and cooling water system, the tank farm, the instrument and service air facilities, the waste water disposal facilities, and the general plant buildings.

5.1.10 Energy Balance

	MM Btu/Hr			
	1000 MW		500 MW	
Energy Consumption	4000 MW System	1000 MW	2000 MW System	500 MW
Dissolver Preheaters	5078	1269	2079	520
Vacuum flash preheater	1291	323	528	132
Wash solvent splitter heater	1176	294	482	120
Ash residue drying	201	50	82	21
Power generation	143	36	59	15
Hydrogen plant fuel	1951	488	799	199
Hydrogen plant feed	3299	825	1350	337
Miscellaneous	201	50	82	21
Total	13,340	3,335	5,461	1,365
Energy Production				
Fuel Gas	9352	2338	3829	957
Light Oil burned as fuel	3988	997	1632	408
Total	13,340	3,335	5,461	1,365

5.2 Complexity

A solvent refined coal (SRC) plant is a complicated process involving many processing steps and many major pieces of equipment. The major processing steps along with the equipment used are listed below:

- Coal Handling and grinding: conveyors, feeders, crushers, screens, dryers, cyclones, bag filters, blowers

- Slurry preheat and dissolvers: gas compressors, expanders, fired heaters, vessels designed for severe process service

- Ash filtering and drying: rotary precoat filters which must remove very small ash particles at high temperature and pressure, blowers, bag filters, compressors, feeders, conveyors, rotary indirect fired dryer, transfer and loading blowers, storage silos

- Solvent, light oil, and cresylic acid recovery: compressors, fractionation towers, fired heaters, steam jets

- Product solidification: flaking drums, conveyors, bucket elevators, storage silos

- Hydrogen plant: fuel fired catalytic reformers

- Sulfur removal and recovery: absorbers, strippers, fired heaters, reactors

- Steam and power generation: boilers, turbines, generators, transformers, electrical switch gear, and electric power distribution facilities

- Other offsites: cooling tower, chemical treatment facilities, air compressors, dryers, filters, waste water treatment facilities, buildings

Also included in almost all sections of the plant are pumps, heat exchangers, tanks, and drums (as well as other processing equipment).

5.3 Flexibility

Due to its complexity an SRC plant is extremely costly and would appear to be economically attractive for supplying power plant fuel only when applied on a very large scale. For example, the use of a centrally located SRC plant which supplied fuel for a number of power plants in the area may be promising (particularly for new power plants). From the results of this study it appears that the minimum economic size for a centrally located SRC plant would be in the range of 30,000-40,000/tpd coal feed (perhaps larger), supplying fuel for about 3,000-4,000/MW of power (at a heat rate of 8700 Btu/kwh and a load factor of 80%).

It is expected that an SRC plant would operate at a relatively constant capacity for 340 days/year (93.2% on-stream factor). The SRC product (liquid or solid) will be stored such that changes in power plant demand for fuel will simply result in a changing inventory of SRC. Long term reduced demand for fuel could be accommodated by shutting down parallel trains of equipment in the various sections of the SRC plant.

Since an SRC plant is so large and complex, start-ups and shut-downs are expected to be rather lengthy and complicated procedures. Therefore, it would appear to be desirable to run the plant at a relatively constant through-put. Major maintenance and inspection work is expected to be done during an annual turn-around of about three weeks duration.

5.4 Status of Technology

5.4.1 Description of Present Status

The solvent refined coal (SRC) process was developed on a laboratory scale by Spencer Chemical Corporation (now Pittsburg and Midway Coal Mining Company). The design basis for the process was developed during the period of 1962-1965 in the pilot plant studies of Spencer Chemical Corporation (1, pp. 17, 19). The pilot plant was designed for 100 lb/hr of feed coal at a 2:1 solvent to coal ratio. The unit is located in Kansas City, Missouri (5, pp. 9, 13).

Process Research, Inc., carried out a conceptual design study for a 222 x 10^9 Btu/day SRC plant which could supply fuel for a 950-1000 MW power station (1, p. 20). A 6 ton/day pilot plant is being built at Southern Electric Company's Ernest C. Gaston Plant. This plant, which is to be used to study the steps in the solvent-refining process, is about ready for start-up. A 50 ton/day pilot plant sponsored by OCR is being built at Tacoma, Washington by Pittsburg and Midway Coal Mining Company. This plant is scheduled for a 1974 start-up (1, p. 17).

Coal News reports that Wheelabrator-Frye Inc., Southern Company, and Gulf Oil Corporation have contracted for construction of a 1000 ton/day SRC demonstration plant. The plant site has not yet been selected. If the demonstration plant is successful, the plan calls for expansion to 10,000 tons/day of SRC. Technology used will be that developed by Pittsburg and Midway Coal Mining Company, a subsidiary of Gulf Oil Corporation, (30, p. 2)

The SRC process is not presently commercial in that no commercial size units are in operation. The process uses unit operations which are commercial. They simply have not as yet been demonstrated for this process.

5.4.2 Areas of Uncertainty

Since the SRC process has never been operated on a commercial scale, there are many areas of uncertainty regarding the process. Some of these are listed below:

- Dissolver temperature. This can vary from 385 to 480°C (725-896°F). The optimum temperature is believed to be about 440°C (825°F). Above this temperature, coking occurs and below it, the viscosity increases rapidly. For example, a ten-fold increase in viscosity of the mixture has been observed as the temperature is decreased from 482°C to 425°C (1, p. 20).

- Solvent to coal ratio. This can vary from 2:1 to 4:1.

- Dissolver residence time. This can vary from 1/4-4 hours.

- Hydrogen requirement. This has been found to vary from 0.8 to 1.5 lb H_2/100 lb of coal feed (1, p. 20).

- Gel formation. Considerable difficulties have been encountered due to the formation of a gel (with an accompanying viscosity increase) as the temperature of the slurry is being increased. The gel disappears when the temperature becomes high enough to form a true solution (1, p. 20; 4, p. 24).

- Phenol and cresylic acid recovery unit. The process for recovery of phenol and cresylic acid from water and hydrocarbon streams is somewhat undefined at the present time. More development work will be necessary to establish a suitable process (2, App. A, p. 3-12).

- Filtration step. The separation of ash from the coal-solvent solution under pressure is expected to be a very troublesome operation. The filtration step must be conducted at 550-700° F and 100-200 psig. The ash solids to be removed are 1-40 microns in size making the filtration task formidable (4, p. 24).

Other methods have been considered for removal of ash from the coal-solvent solution. Some of these are listed below:

- Centrifuges
- Hydroclones
- Cartridge filters

- Degree of ash removal. Ash in the solvent refined coal is expected to range from 0.05 to 0.10% (1, p. 20; 2, App. B, p. 3-1). Pilot plant studies have shown that ash in the solvent refined coal varies from 0.17 to 0.48% (5, pp. 252-253).

- Degree of sulfur removal. This will vary and will depend on the type of sulfur originally present in the coal as well as on the process operating conditions (3, p. 207). The process will remove virtually all of the pyritic sulfur and about 50-70% of the organic sulfur originally present (4, p. 24). Pilot plant studies have shown that the solvent refined coal has a sulfur content of 0.45 to 1.22% when the feed coal contained 0.81 to 4.18% sulfur on an as received basis (5, pp. 252-253).

- Product distribution. Of particular importance is the amount of cresylic acid produced. This can vary from 1-4% of the coal feed. For this study a value of 1.2% was used (2, App. A, p. 4-4). Also important is the quantity of phenol produced. This can vary from 0.2-0.5% of the coal feed. A value of 0.36% was used for this study (2, App. A, p. 4-4).

5.5 Environmental Effects

The primary purpose of an SRC plant as evaluated in this study is to produce a low sulfur, ash-free fuel for power plant use. The sulfur content of the SRC is expected to be about 1% (2, p.10; 3, p. 206) which would result in a power plant emission of 1.25 lb SO_2/MM Btu assuming an SRC higher heating value of 16,000 Btu/lb. This almost exactly matches the current EPA emission standard of 1.20 lb SO_2/MM Btu. If emission standards become more stringent, the sulfur content of the SRC can probably be reduced to 0.4-0.6% by process modifications at increased cost (1, p. 23; 3, p. 207). A sulfur level of 0.6% in the SRC would result in a power plant emission of 0.75 lb SO_2/MM Btu.

Elemental sulfur is formed in the SRC plant via a conventional Claus unit after the H_2S-rich gases have been treated in an amine unit. The quantity of sulfur produced is shown below for the different plant sizes.

Power Plant Size	1000 MW		500 MW	
SRC Plant Size	4000 MW	1000 MW	2000 MW	500 MW
Sulfur Product: LT/D	861	215	352	88.1

In the economic study, the sulfur was given a value of $5/LT which amounts to a credit of 0.62 ¢/MM Btu. This is virtually negligible in the overall cost of the process (<1%). The economics would not be significantly affected if the sulfur had no credit at all. If a significant increase in sulfur selling price could be acheived (say to $30/LT), the process cost would drop by about 3.10 ¢/MM Btu.

There appears to be some question about the ash content of SRC. Some sources expect it to be as low as 0.05% (1, p. 20) while others predict a value of 0.10% (2, App. B, p. 3-1).

Pilot plant studies have yielded somewhat higher values-0.17 to 0.48% (5, pp. 252 - 253). If the SRC is assumed to have 0.20% ash, the overall ash removal efficiency of the process will be about 99%, assuming that the coal originally contains about 10-12% ash.

$$\% \text{ Ash Removal} = 100 - \left[\frac{0.576\ (0.20)}{11.5} \right] 100 = 99.0$$

The SRC process was modified to eliminate the carbon burn-off section of the plant. Therefore, the ash from the drying section will contain about 30-35% carbon (2, App. A, p. 5-21). This increases the quantity of material to be disposed of by 50%. It is assumed that the ash (containing carbon) will be used as land-fill.

Substantial quantities of cresylic acids are formed in the SRC process. These compounds reportedly can vary from 1-4% of the coal feed. For this study, a value of 1.2% was used (2, App. A, p. 4-4). The following table shows the amount of cresylic acid assumed to be produced for the different plant sizes.

Power Plant Size	1000 MW		500 MW	
SRC Plant Size	4000 MW	1000 MW	2000 MW	500 MW
Cresylic Acid: T/D	488	122	200	49.9

A sales price of \$100/T (5¢/lb) was used for cresylic acid. The sale of cresylic acid has a significant impact on the process economics. If either the sales price or the amount produced doubled, the process operating cost would drop by 7.07 ¢/MM Btu.

Carbon dioxide is produced in large volumes in the hydrogen plant. No credit was taken for this compound.

There are several waste water streams from the process. Process water which contains cresylic acid will be filtered through activated carbon filters. Process area runoff water will go to a separate storm sewer and to a holding pond. Oil separation will be accomplished by an API skimmer. Aeration may be required. Other area runoff water and cooling tower blowdown will be discharged to the sewer with no treatment. Sewage will be treated in a package unit before being discharged (2, App. A, p. 3-14).

5.6 Installation

Since the SRC process is not commercial at the present time, some uncertainties exist when attempting to predict the length of time required for a full size unit to be placed on stream.

As shown in Figure 5.2, about 11-12 years are believed to be required for commercialization of a large SRC process. The time can be broken down briefly as follows:

- Demonstration plant: 3 years
- Feasibility studies, environmental studies, licenses and permits, design and engineering: 4 years
- Procurement and construction: 4 years
- Start-up: 1 year

An SRC plant is a complex process employing many processing steps and many pieces of equipment. In addition, the process appears to be economic only when applied at a very large scale. Although no layout drawings are available at the present time, it is believed that a plant capable of processing 10,000 T/D of feed coal would require about 50 acres of land - perhaps more if additional raw coal storage is to be provided for. Naturally, larger plants will require even more space.

As stated previously, the installation of a large, centrally located SRC plant appears to be the most promising. The plant would be sized for about 30,000 - 40,000 T/D of coal feed and would supply fuel for about 3,000 - 4,000 MW of power. Presumably, the power would be generated by several stations and the SRC would be transported to the power plants as required by rail or barge.

FIGURE 5.1

SOLVENT REFINED COAL PROCESS FLOW DIAGRAM

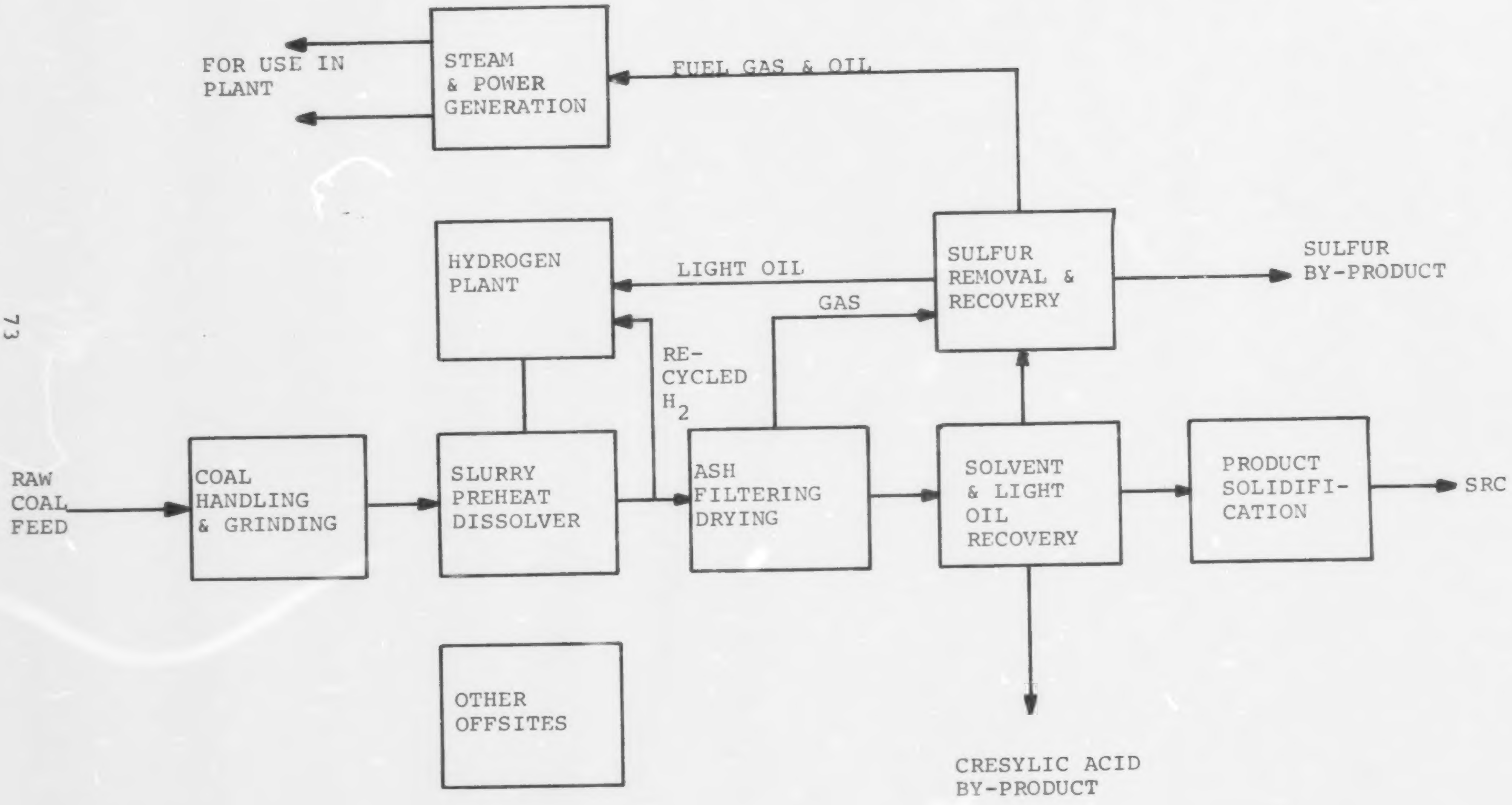
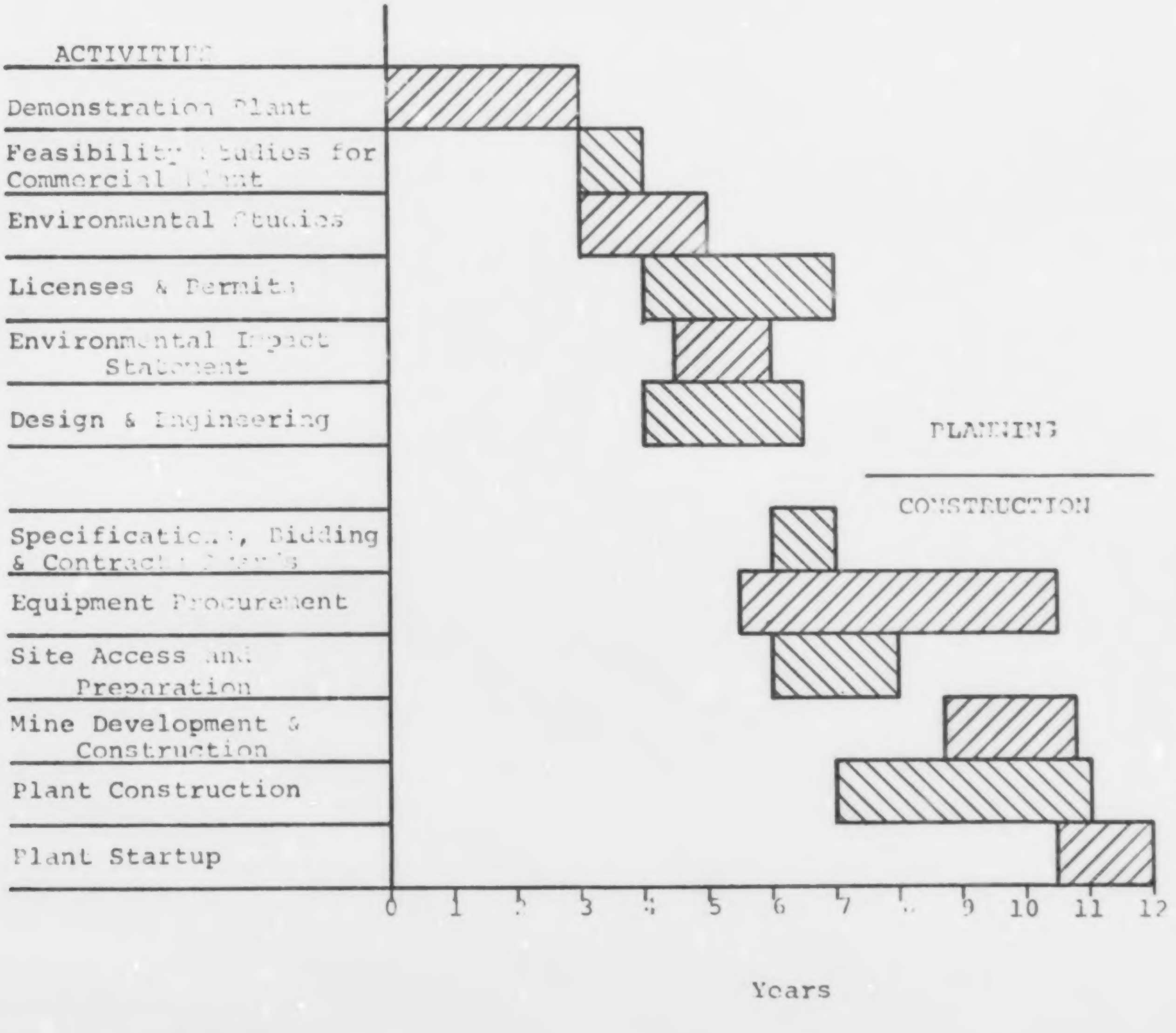

STEAM & POWER GENERATION
FOR USE IN PLANT
FUEL GAS & OIL
HYDROGEN PLANT
LIGHT OIL
GAS
SULFUR REMOVAL & RECOVERY
SULFUR BY-PRODUCT
RE-CYCLED H_2
RAW COAL FEED
COAL HANDLING & GRINDING
SLURRY PREHEAT DISSOLVER
ASH FILTERING DRYING
SOLVENT & LIGHT OIL RECOVERY
PRODUCT SOLIDIFI-CATION
SRC
OTHER OFFSITES
CRESYLIC ACID BY-PRODUCT

Fig. 5.2

ANTICIPATED COAL LIQUEFACTION PROJECT SCHEDULE

(REF. 8)



6. LOW BTU GAS

6.1 Process Description

Supplies of low sulfur petroleum fuels are becoming scarce, therefore, power plants will have to use high sulfur fuel but with some kind of desulfurization. Production of intermediate or low Btu gas from coal, which can be used in a power plant, is a possible alternative. This route has two major advantages:

1) the gas can be generated under pressure which is an advantage for the existing as well as the new power plants since it permits some operating economics.

2) the gas is produced under reducing conditions which convert coal sulfur to H_2S which is readily removed by well-proven absorption processes.

Types of Gasifiers

There are three types of gasifiers that are being investigated on various levels of development. They are:

1) Fixed Bed Gasifiers: This type has the advantage of a counter-current flow which aids the overall conversion considerably. It also has the advantage of having distinct temperature zones inside the gasifier. However, it has some disadvantages such as the low throughput and the restriction of feed to noncaking pre-sized coal particles. This type of gasifier is best exemplified by the Lurgi gasifier (7).

2) Entrained Bed Gasifiers: This type has the advantage of high throughput, and it can use any type of coal without requiring sized coal particles. It has

the disadvantages of high temperature operation and the requirement for large volume reactors since complete conversion is difficult to acheive in one pass. This type is best exemplified by Koppers-Totzek gasifier (7).

3) Fluidized Bed Gasifiers: The major advantages of this type are the long solids residence time and the excellent heat transfer characteristics which aid in char gasification. It has some disadvantages in that it has to operate below ash fusion temperature to prevent agglomerates formation and it also requires some pretreatment for caking coals. This type is exemplified by the Hygas gasifier (7).

Since the Lurgi gasifier is the most advanced type (62 gasifiers are already installed in 14 installations) it will be used for the purposes of this comparative study.

The Lurgi Gasifier

The gasifier is the major piece of equipment in a Lurgi gasification plant. Coal is fed through a lock hopper. In the gasification zone, coal is contacted with a mixture of steam and air. The coal is gasified and a stirring mechanism keeps the coal and ash particles from fusing together. The ash is then recovered at the bottom through a revolving grate into a lock hopper from which it is ultimately discharged. The raw gas leaving the gasifier contains liquid hydrocarbons and tars distilled from the coal. They are condensed and are either recycled to the gasifier or used as fuel gas. This leaves a gas containing H_2S which is removed by conventional purification equipment.

For this study, a Benfield system is employed for H_2S removal. The concentrated H_2S stream is converted to elemental sulfur in a conventional 2-stage Claus plant. The Claus plant tail gas is incinerated and vented to atmosphere; no tail gas treatment is included. The system has been designed to give an overall sulfur recovery of 90%. Total emissions from the gasification plant and power plant are 0.6 lbs SO_2/MM Btu, based on the coal feed to the gasifiers.

Figure 6.1 shows a schematic diagram for producing low Btu gas by the Lurgi process. For the purpose of this investigation two gasification plants were considered. The first supplies sufficient energy to a 500 MW existing power plant operating at a load factor of 60%, and the second supplies a new 1000 MW power plant operating at 80% load factor. Both plants utilize coal having 3.5% by weight sulfur, 5% by weight moisture and 12% by weight ash. The heating value for coal was assumed to be 12,000 Btu/lb. The heat rate for the existing plant was assumed to be 9,500 Btu/kwh while that for the new plant was assumed to be 8,700 Btu/kwh. Table 6.1 shows the material balance for both cases, while Table 6.2 presents the annual utilities consumption.

6.2 Complexity

The major process steps are shown in the flow diagram of Figure 6.1. The gasification section for a 1000 MW power plant consists of thirty trains of gasifiers with their accessories. There are three trains of equipment for fines agglomeration, H_2S removal, and sulfur recovery. The Claus plant is a two-stage system with no tail gas clean up. Although a three-stage Claus plant may cut SO_2 emissions from the sulfur plant in half, it was assumed that the added investment would not be justified since the sulfur plant emissions are below regulation standards (see Table 6.1). In each section of this process, several heat exchangers, pumps and columns are involved. The gasifier design is the only complex portion of this plant. The design of the rest of the equipment is relatively simple.

6.3 Flexibility

Operation of a coal gasification plant is more like that of a chemical plant than of a power plant. It is relatively complex since it requires contorl of the flow and composition of more process streams. It is more difficult to start up and shut down and it requires more time to reach optimum conditions than does a power plant. Because of these reasons, special training of operators may be necessary and could prove to be expensive for small power plants. Another major disadvantage is the necessity to start up and shut down the facility simultaneously with the power plant since large-scale storage of gas is very costly and impractical. For example, for a 500 MW existing plant operating at 60% load factor it would require about 250,000 brake horsepower to compress only one hour's gas production from 16 psi to 500 psi in a three stage compressor with cooling between stages. The compressed gas needs a spherical storage vessel of 130 feet inside diameter.

Production of low Btu gas could therefore be more applicable to industrial power plants than to utility plants because the output of such a plant does not vary appreciably and thus the gasification plant can be designed for a relatively constant output which avoids the problems of instability during frequent start ups and shut downs or during cyclic increases and decreases in production.

6.4 Status of Demonstrated Technology

Several gasification processes are being investigated at various levels of development. Some of these are already into the commercial stage. Table 6.3 shows these various projects and their stage of development.

6.5 Environmental Effects

The major effect on the environment is the disposal of ash which amounts to about 200 M tons/yr for the 500 MW plant and approximately 500 M tons/yr for the 1000 MW plant. This ash can be disposed of by land filling. The sulfur produced from the 500 MW plant would amount to 52 M tons/yr, and that from the 1000 Mw plant about 126 M tons/yr. It is assumed that such amounts of sulfur can be readily marketed and should not present any major problem. SO_2 emissions would amount to approximately 0.58 lbs SO_2/MM Btu of coal gasified, which is below the national standards of 1.2 lbs SO_2/MM Btu.

A major environmental advantage of the low Btu gas combustion in power plants is the reduction of NO_x emissions by 60 - 90% over the conventional boilers (11).

6.6 Installation

As shown in Figure 6.2, it requires eight to nine years to get the gasification plant to a producing stage. About half this time is needed for preliminary studies and design of the facility. The second half is spent in construction and start up of the plant. The Lurgi gasification plant requires a relatively large piece of land since several trains of gasifiers are needed. It was assumed for this study that the gasification plant could be located adjacent to the power plant thereby minimizing the length of gas transmission lines.

TABLE 6.1

MATERIAL BALANCE FOR LOW BTU GAS PROCESS*

STREAM	EXISTING 500 MW POWER PLANT - 60% LF	NEW 1000 MW POWER PLANT - 80% LF
Raw Coal Feed: T/HR	314	575
(3.5% S; 12% Ash; 5% H_2O) MM BTU/HR	7,536	13,800
Steam: T/HR	240.1	439.8
Ash: T/HR	37.6	69
Air: MM SCF/HR	15.2	27.8
Sulfur: T/HR**	9.9	18.1
Clean Gas: MM SCF/HR	38.67	70.82
: 10^9 BTU/HR	4.75	8.7
SO_2 Emissions, lbs. SO_2/MM BTU coal feed***	0.58	0.58

* Reference: "Clean Fuel Gas From Coal", Lurgi Mineraloltechnik GmbH, October, 1971.

** 90% recovery

*** Total emissions for both gasification plant and power plant.

TABLE 6.2

ANNUAL UTILITIES - LOW BTU GAS PLANTS

ITEM	EXISTING 500 MW PLANT		NEW 1000 MW PLANT	
	Quantity	Cost M$	Quantity	Cost M$
Feed Water \$0.20/Mgal.	1.29×10^5 Mgal	25.8	3.15×10^5	63.0
Cooling Water \$0.02/Mgal.	230.2×10^5 Mgal	460.4	562.7×10^5	1125.5
Steam \$0.50/M lb	25.2×10^5 Mlb	1261.9	61.6×10^5	3082.1

Reference: "Clean Fuel Gas From Coal", Lurgi Mineraloltechnik, October, 1971.

TABLE 6.3

COAL GASIFICATION PROCESSES FOR PRODUCTION OF LOW B.T.U. GAS

PROCESS	REACTOR BED TYPE	NATURE OF RESIDUE	PRESSURE ATM.	TEMP. °F	CAPACITY T/D
Commercial					
Winkler	Entrained	Dry Ash	1	1500	2000
Demonstration					
Lurji	Fixed	Dry Ash	20	1000	2000
Pilot Plants					
1. USBM	Fixed	Dry Ash	20	1000	20
2. G.E.	Fixed	Dry Ash	8	1000	0.25
3. Combustion Engineering	Entrained	Slag	1	>2100	180
Consolidated Edison					
4. Westinghouse	Multiple Fluid Beds	Dry Ash	10-16	1300-1700 & 2000	15
5. Pittsburg-Midway	Entrained (2-stage)	Slag	4-35	>2100	1200
6. IGT-U-gas	Fluid Bed	Dry Ash	20	1900	30-50

SOURCE: - Chemical Engineering, July 22, 1974

FIGURE 6.1

LURGI LOW BTU GAS PROCESS

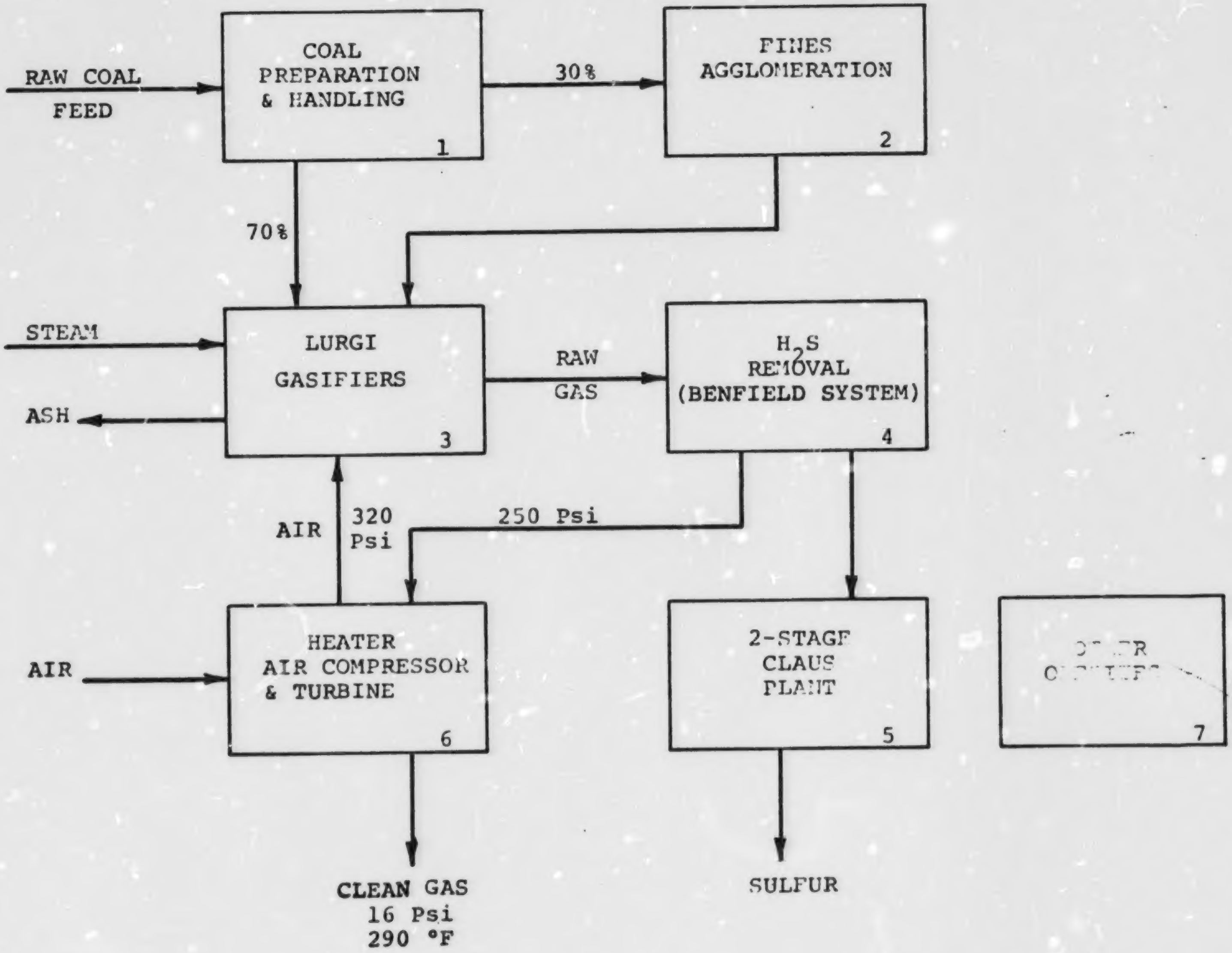
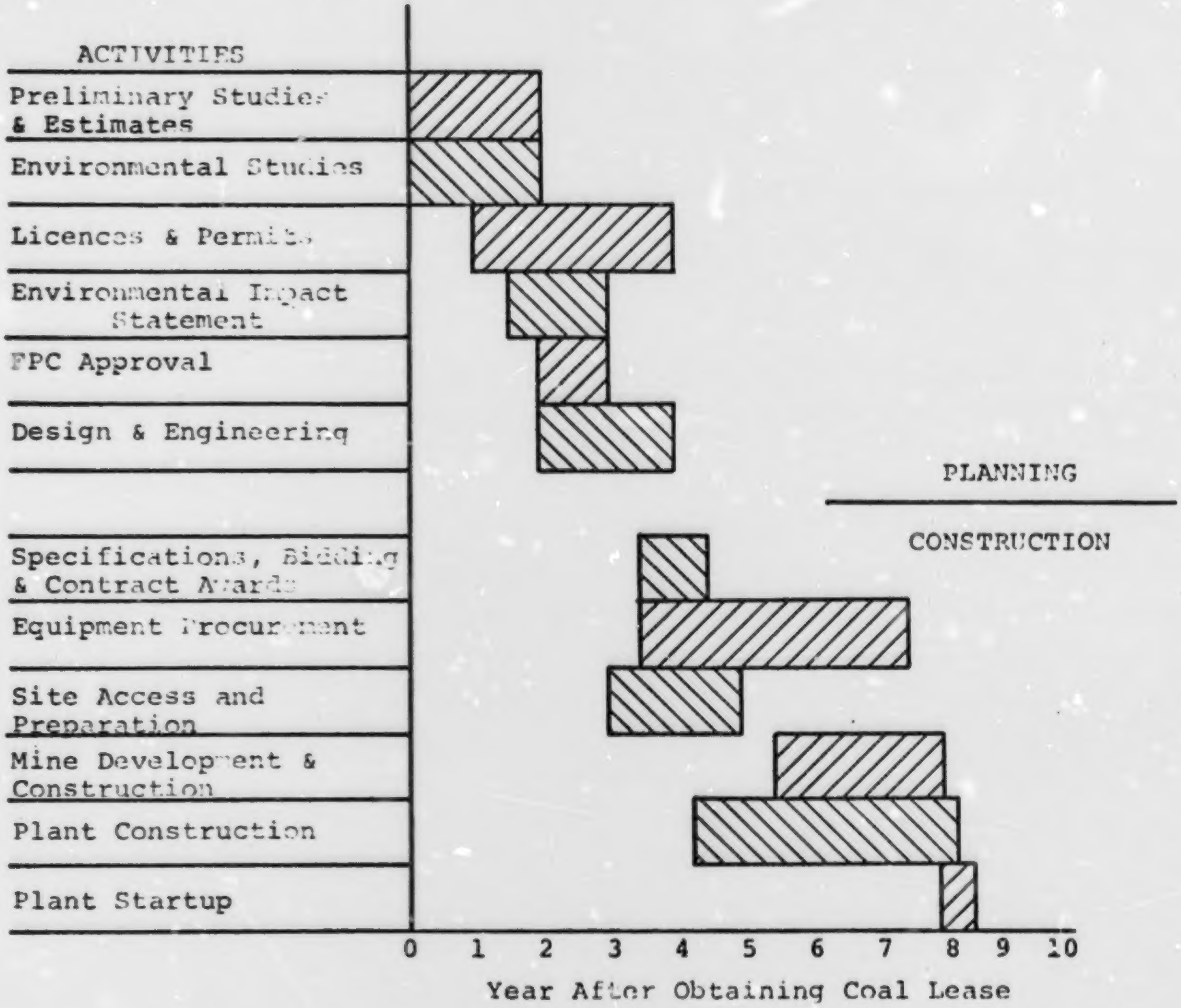

RAW COAL
FEED
COAL PREPARATION & HANDLING
1
30%
FINES AGGLOMERATION
2
70%
STEAM
ASH
LURGI GASIFIERS
3
RAW GAS
H_2S REMOVAL (BENFIELD SYSTEM)
4
AIR
320 Psi
250 Psi
AIR
HEATER AIR COMPRESSOR & TURBINE
6
2-STAGE CLAUS PLANT
5
7
CLEAN GAS
16 Psi
290 °F
SULFUR

Fig. 6.2

ANTICIPATED COAL GASIFICATION PROJECT SCHEDULE

(REF. 8)



7. ECONOMIC COMPARISON OF PROCESSES

7.1 Basis for Costs

Investment and operating costs for the wet limestone process, the Wellman/Allied process, and the solvent refined coal process were obtained using cost models developed by Kellogg and reported in a recent study for EPA (7). The wet limestone model is based on a study by Catalytic, Inc. (31). The basis for the Wellman Lord/Allied model is the demonstration plant now under construction at the D. H. Mitchell plant of the Northern Indiana Public Service Company. The cost model for solvent refined coal is based on the report by Stearns-Roger for the Pittsburg and Midway Coal Mining Comapny (2). Costs for the Cat-Ox process were taken from a recent TVA study for EPA (12). These were adjusted to a basis consistent with the other processes. Cost data for Lurgi low Btu gas were taken in part from a cost model for substitute natural gas, as reported in the Kellogg study previously cited, and partially from confidential Kellogg sources.

To obtain a consistent economic basis for the processes so that valid cost comparisons could be made, a standard cost accounting format has been utilized. This format, which is based on a utility financing method recommended in a report to the Federal Power Commission (32), was used by Kellogg in the report previously mentioned to develop a general cost model. The general cost model has been used for all cost comparisons. A detailed description of the model including definition of terms is discussed in the Appendix.

All plants are assumed to have a midwest (Cincinnati) location. Costs are on an end-of-1973 basis, since most of the available cost information is referenced to this time. No contingencies are included in the estimates.

To further establish a uniform cost basis, standard unit prices were assumed for all raw material, utilities, and byproducts. These values are listed in Table 7.1.

7.2 Energy Conversion Efficiency

Table 7.2 presents a summary of the energy conversion efficiencies and heat rates of each process. These are based on the data developed in preceding sections of the report on the energy consumption of each process. The low Btu gasification process is the least efficient followed by SRC and then stack gas scrubbing.

7.3 Manpower Requirements

Table 7.3 presents the total number of operators for each process. It is clear that stack gas scrubbing requires far less working labor than the other two processes.

7.4 Economics of Each Process

7.4.1 Flue Gas Desulfurization

This section details an economic analysis of three flue gas desulfurization processes: wet limestone, Wellman-Lord/Allied, and Cat-Ox. Tables 7.4, 7.5, and 7.6 outline the annual scrubbing costs of the three processes. Figures 7.1, 7.2, and 7.3 add the scrubbing cost to the coal cost, and show the total production cost of energy to the power plant in \$/MM Btu. Figure 7.4 shows the effect of varying load factor on the total production cost.

The wet limestone and Wellman-Lord/Allied scrubbing costs are based on the M. W. Kellogg scrubber cost models (3, pp. 77 - 135). The Cat-Ox scrubbing costs are calculated from equipment and utility costs outlined in the TVA report (12, pp 244 - 249), and from equations presented in the M. W. Kellogg general cost model (3 pp 54 - 76).

In the Wellman-Lord/Allied and Cat-Ox processes, credits are given for the by-products generated.

7.4.2 Economics of SRC

The total plant investment (TPI) for the SRC process was calculated based on the MWK cost model (3, pp. 198 - 203). This model in turn was based on the Stearns-Roger design (2, App. A). For the 1000 MW new plant, a credit of \$30/kw was applied to the TPI for the SRC process due to savings at the power plant (coal handling, ESP's, ash handling, etc.).

Total capital required (TCR) includes interest during construction, startup costs, and working capital and is calculated by the following:

$$TCR = 1.21\ TPI + 0.8\ (TO)\ (CO)\ (1 + F) + 0.4\ ANR$$

Where TO = Total number of shift operators

CO = hourly rate per operator (Gulf Coast), \$1 hr.

F = Location Factor (=1.53 for Cincinnati)

ANR = Annual cost of raw materials and utilities less by-product credits, M\$/yr.

For the SRC process, ANR is calculated from

$$ANR = ACOAL + ACHEM - ASULF - ACRES$$

Where ACOAL = Annual cost of coal, M$/yr
ACHEM = Annual cost of chemicals, M$/yr
ASULF = Annual credit for sulfur, M$/yr
ACRES = Annual credit for cresylic acid, M$/yr

The total annual production cost (TAC) including return of capital, payment of interest, and income tax on equity return is given by:

$$TAC = 0.225\ TPI + 2.1\ (TO)\ (CO)\ (1 + F) + 1.04\ ANR$$

The total annual production cost for the SRC process was calculated in M $/yr and ¢/MM Btu for four different cases. The effect of coal price on TAC was investigated (coal prices of $5, $10, and $15/T were used). Also the effect of load factor on TAC (at a coal price of $10/T) was calculated.

7.4.3 Economics of Low Btu Gas

The following is a detailed discussion of the cost figures appearing in Tables 7.11 - 7.12 which display a summary of the process economics.

Total Plant Investment

The bare costs (BARC) of sections 1, 2, 3, 4 and 7 were determined using a cost estimation model for a base case of an SNG plant developed by MWK. Some adjustment was made in the cost of Section 7 since the cost of this section would be higher for the SNG plant. The bare costs of Section 5 and 6 were obtained by proper adjustments of current prices offered by some producing companies at the end of 1973 (9, 10).

As shown in the Appendix, the plant investment (TPI) can be approximated as 1.12 BARC, where BARC is the sum of equipment and other material costs, labor costs, and home office engineering costs. This relation was used in the calculation of the individual section's investment.

For comparison of low Btu gas vs. stack gas scrubbing to service a new power plant, it was estimated that the low Btu gas plant can be credited with 40 \$/kw of power plant capacity due to savings in the power plant investment. Such savings are a result of eliminating some major equipment from the power plant (such as coal handling, ESP units, etc.) when it uses clean gas as a fuel.

Annual Cost of Raw Material and Utilities less Credits (ANR)

ANR can be calculated from:

ANR = Annual Coal Cost (ACOAL) + Annual Utilities - Annual Credits

ANR was calculated for three prices of the coal feed as shown in Tables 7.11 - 7.12. Changes in the steam cost were considered to be small in comparison with those of the changes in the coal price, and thus were neglected.

Total Capital Required

The total capital required (TCR) for the gasification plant is:

$$TCR = 1.15\ TPI + 0.8\ (TO)(CO)\ (1 + F) + 0.4\ ANR$$

This equation was used in the calculations. Two values for TCR of the new plant are shown in Table 7.12 as a result of using the adjusted TPI which takes into account the 40 \$/kw credit for comparison with stack gas scrubbing.

Total Annual Production Cost

Based on a twenty year life for the gasification plant, the total annual cost of production is:

$$TAC = 0.217\ TPI + 2.1\ (TO)(CO)\ (1 + F) + 1.04\ ANR$$

The changing price of coal will affect only the last term of the above equation.

Values for total annual production cost on a yearly basis as well as on a cents per MM Btu basis are shown in Tables 7.11 - 7.12 for three different coal prices. The effect of load factor on TAC for an existing plant and a new plant is shown in figure 7.7, for a coal price of $10/Ton. The effect of coal cost on TAC is presented for these two plants in figure 7.8.

Doubling the coal price form 5 to 10 $/T results in a 32% increase in the cost of producing the gas and the effect is linear in the cost range 5 - 15 $/T.

7.5 Cost Comparison

Figures 7.9 - 7.10 show the total annual cost of all the investigated processes vs. coal cost for the existing 500 MW plant operating at 60% load factor and the new 1000 MW plant operating at 80% load factor respectively. For both plants, stack gas scrubbing appears to be superior to the use of SRC or low Btu gas. The wet limestone process semms to be the cheapest scrubbing method and costs only about half that of SRC or low Btu gas. Quadrupling the size of the SRC plant reduces the cost of production by almost 40% and makes the process competetive with stack gas scrubbing at low coal prices.

If the coal costs less than 10 $/ton, low Btu gas is cheaper than SRC if the plants are to service only one 500 MW existing or 1000 MW new power plant.

TABLE 7.1

UNIT PRICES USED IN COST COMPARISONS

(Basis: End of 1973)

Item	Unit Price
Limestone	$ 4.00/ton
Ammonia	$ 50.00/ton
Sodium Carbonate	$ 40.00/ton
Filter-Aid	$ 50.00/ton
Catalyst (Cat-Ox)	$ 44.00/cu. ft.
Steam	$ 0.50/M lbs
Process Water	$ 0.20/M gal
Cooling Water	$ 0.02/M gal
Power	8.0 mills/kwh
Natural Gas	$ 0.50/MSCF
Fuel Oil	$ 0.80/MM Btu
Sulfur	$ 5.00/LT
Cresylic Acid	$100.00/ton
Sulfuric Acid (80%)	$ 4.63/ton

Table 7.2

Process Energy Conversion Efficiency

Power Plant	Power Plant with no SO_2 controls	Flue Gas Desulfurization			SRC	Low Btu Gas
		Wet Limestone	Wellman-Lord/Allied	Cat-Ox		
EXISTING 500 MW						
% Efficiency	35.9	33.5	30.6	35.1	26.1	22.6
Heat Rate BTU/KWH	9,500	10,184	11,139	9,717	13,060	16,132
NEW 1000 MW						
% Efficiency	39.2	37.7	33.6	39.2	28.5	24.7
Heat Rate BTU/KWH	8,700	9,047	10,157	8,700	11,960	14,770

Table 7.3

Process Manpower Requirements*

Power Plant	Flue Gas Desulfurization			SRC		
	Wet Limestone	Wellman-Lord/Allied	Cat-Ox	Sized For Power Plant	Sized for 4 Times Larger	Low Btu Gas
EXISTING 500 MW						
Total Operators	8	16	8	90	222	159
NEW 1000 MW						
Total Operators	8	16	8	160	396	300

* References: 3,12

Table 7.4

ECONOMICS OF WET LIMESTONE SCRUBBING

	Existing 500 MW	New 1000 MW
Total Plant Investment, M$	29,115	40,799
Total Capital Requirement, M$	34,880	48,610
Raw Material & Utility Cost, M$/yr		
Limestone	712.5	1,740
Ammonia	10.6	19.5
Process Water	27.2	66.5
Fuel Oil	283.5	692.4
Electricity	586.4	1,432.0
ANR, M$/yr	1,620.2	3,950.4
Total Annual Scrubbing Costs		
M$/YR	9,029	14,153
MILLS/KWH	3.43	2.01
¢/MM BTU	36.11	23.09

Table 7.5

ECONOMICS OF WELLMAN-LORD/ALLIED SCRUBBING

	Existing 500 MW	New 1000 MW
Total Plant Investment, M$	31,921	46,561
Total Capital Requirement, M$	37,820	55,930
Raw Material and Utility Costs Less Credits, M$/yr		
Sodium Carbonate	334.9	817.8
Natural Gas	216.7	529.3
Filter Aid	18.3	44.7
Electricity	551.5	1,346.8
Steam	806.0	1,968.4
Cooling Water	123.2	300.9
Process Water	3.8	9.3
Fuel Oil	283.5	692.4
Sulfur Credit	(141.6)	(345.8)
Purge Disposal	16.6	40.6
ANR, M$/yr	2,212.9	5,404.2
Total Annual Scrubbing Costs		
M$/YR	10,477	17,259
MILLS/KWH	3.98	2.46
¢/MM BTU	41.90	28.29

Table 7.6

ECONOMICS OF CAT-OX SCRUBBING

	Existing 500 MW	New 1000 MW
Total Plant Investment, M$	42,572	78,534
Total Capital Requirement, M$	49,498	90,435
Raw Material & Utility Costs Less Credits, M$/yr		
Catalyst	138.4	351.5
Electricity	470.4	1,495.9
Fuel Oil	960.0	-
Cooling Water	133.7	13.6
Steam Credit	-	(640.0)
Sulfuric Acid Credit	(636.1)	(1,203)
ANR, M$/yr	1,066.4	18.0
Total Annual Scrubbing Cost		
M$/YR	11,496	18,929
MILLS/KWH	4.37	2.70
¢/MM BTU	46.05	31.05

Table 7.7

SRC PROCESS — 80% LF

1000 MW NEW POWER PLANT — 60.97×10^{12} BTU/YR

1000 MW SRC PLANT

Total Plant Investment

Section	Description	Cost: M$
1	Coal Preparation	10,000
2	Preheater, Dissolvers	40,000
3	Ash Filtration, Drying, Disposal	15,000
4	Solvent, Light Oil, Cresylic Acid Recovery	30,000
5	Product Solidification, Handling, Storage	10,000
6	Hydrogen Plant	10,000
7	Sulfur Removal and Recovery	10,000
8	Steam and Power Generation	10,000
9	Other Offsites	29,000
TPI		164,000
Credit		30,000
TPI Adjusted		134,000

	Coal Price		
	$5/T	$10/T	$15/T
TCR: M$	206,000	212,000	219,000
TCR Adjusted: M$	169,700	175,700	182,700
Raw Materials and Utilities Costs Less Credits M$/YR			
ACOAL	16,582	33,163	49,745
ASULFUR: $5/LT		(366)	
ACRESYLIC ACID: $100/T		(4,145)	
ACHEMICALS		359	
ANR: M$/YR	12,430	29,011	45,593
Total Annual Production Cost			
0.225 TPI		30,150	
2.1 (TO) (CO) (1 + F)		5,950	
1.04 ANR	12,926	30,171	47,417
TAC: M$/YR	49,026	66,271	83,517
: ¢/MM BTU	80.4	108.7	137.0

Table 7.8

SRC PROCESS

1000 MW NEW POWER PLANT

4000 MW SRC PLANT

80% LF

243.8×10^{12} BTU/YR

Total Plant Investment

Section	Description	Cost: M$
1	Coal Preparation	24,400
2	Preheater, Dissolvers	97,700
3	Ash Filtration, Drying, Disposal	36,700
4	Solvent, Light Oil, Cresylic Acid Recovery	73,300
5	Product Solidification, Handling, Storage	24,400
6	Hydrogen Plant	24,400
7	Sulfur Removal and Recovery	24,400
8	Steam and Power Generation	24,400
9	Other Offsites	73,300
TPI		403,000
Credit		120,000
TPI Adjusted		283,000

	Coal Price		
	$5/T	$10/T	$15/T
TCR: M$	513,100	539,700	566,200
TCR Adjusted: M$	367,900	394,500	421,000
Raw Materials and Utilities Costs Less Credits, M$/YR			
ACOAL	66,326	132,652	198,978
ASULFUR: $5/LT		(1,464)	
ACRESYLIC ACID: $100/T		(16,580)	
ACHEMICALS		1,436	
ANR: M$/YR	49,718	116,044	182,370
Total Annual Production Cost			
0.225 TPI		63,675	
2.1 (TO) (CO) (1 + F)		14,728	
1.04 ANR	51,707	120,686	189,665
TAC: M$/YR	130,110	199,089	268,068
: ¢/MM BTU	53.4	81.7	110.0

Table 7.9

SRC PROCESS 60% LF

500 MW EXISTING POWER PLANT 24.97 x 10^{12} BTU/YR

500 MW SRC PLANT

Total Plant Investment

Section	Description	Cost: M$
1	Coal Preparation	5,600
2	Preheater, Dissolvers	22,200
3	Ash Filtration, Drying, Disposal	8,300
4	Solvent, Light Oil, Cresylic Acid Recovery	16,600
5	Product Solidification, Handling, Storage	5,600
6	Hydrogen Plant	5,500
7	Sulfur Removal and Recovery	5,600
8	Steam and Power Generation	5,500
9	Other Offsites	16,600
TPI		91,500
Credit		--
TPI Adjusted		91,500

	Coal Price		
	$5/T	$10/T	$15/T
TCR: M$	114,000	116,700	119,500
TCR Adjusted: M$	114,000	116,700	119,500
Raw Materials and Utilities Costs Less Credits, M$/YR			
ACOAL	6,788	13,576	20,364
ASULFUR: $5/LT		(150)	
ACRESYLIC ACID: $100/T		(1,697)	
ACHEMICALS		147	
ANR: M$/YR	5,088	11,876	18,664
Total Annual Production Costs			
0.225 TPI		20,588	
2.1 (TO) (CO) (1 + F)		3,347	
1.04 ANR	5,292	12,351	19,411
TAC: M$/YR	29,227	36,286	43,346
: ¢/MM BTU	117.0	145.3	173.6

Table 7.10

SRC PROCESS

60% LF

500 MW EXISTING POWER PLANT

99.86×10^{12} BTU/YR

2000 MW SRC PLANT

Total Plant Investment

Section	Description	Cost: M$
1	Coal Preparation	13,600
2	Preheater, Dissolvers	54,600
3	Ash Filtration, Drying, Disposal	20,600
4	Solvent, Light Oil, Cresylic Acid Recovery	40,900
5	Product Solidification, Handling, Storage	13,600
6	Hydrogen Plant	13,600
7	Sulfur Removal and Recovery	13,600
8	Steam and Power Generation	13,600
9	Other Offsites	40,900
TPI		225,000
Credit		--
TPI Adjusted		225,000

	Coal Price		
	$5/T	$10/T	$15/T
TCR: M$	283,500	294,400	305,300
TCR Adjusted: M$	283,500	294,400	305,300
Raw Materials and Utilities Costs Less Credits, M$/YR			
ACOAL	27,166	54,332	81,498
ASULFUR: $5/LT		(599)	
ACRESYLIC ACID: $100/T		(6,792)	
ACHEMICALS		588	
ANR: M$/YR	20,363	47,529	74,695
Total Annual Production Cost			
0.225 TPI		50,625	
2.1 (TO) (CO) (1 + F)		8,256	
1.04 ANR	21,177	49,430	77,683
TAC: M$/YR	80,058	108,311	136,564
: ¢/MM BTU	80.2	108.5	136.8

Table 7.11

LOW BTU GAS PROCESS

500 MW EXISTING POWER PLANT

(60% LF - 24.97 x 10^{12} BTU/YR.)

Total Plant Investment

Section	Description	Cost: M$
1	Coal Preparation % Handling	2,450
2	Fines Agglomeration	5,794
3	Coal Gasification	16,710
4	Claus Plant	2,663
5	H_2S Removal Unit	2,263
6	Heater, Air Compressor and Turbine	4,723
7	Other Offsites	15,611
TPI No Credit		50,214

	Coal Price		
	$5/T	$10/T	$15/T
TCR: M$	63,211	66,512	69,813
Raw Materials and Utilities Costs Less Credits, M$/YR			
ACOAL	8,252	16,504	24,756
ASULFUR: $5/LT		(232)	
UTILITIES		1,748	
ANR: M$/YR	9,768	18,020	26,272
Total Annual Production Cost			
0.217 TPI		10,896	
2.1 (TO) (CO) (1 + F)		5,950	
1.04 ANR	10,159	18,741	27,323
TAC: M$/YR	27,005	35,587	44,169
: ¢/MM BTU	108.1	142.5	176.9

Table 7.12

LOW BTU GAS PROCESS

1000 MW NEW POWER PLANT

(80% LF - 60.97 x 10^{12} BTU/YR.)

Total Plant Investment

Section	Description	Cost: M$
1	Coal Preparation & Handling	3,523
2	Fines Agglomeration	7,381
3	Coal Gasification	34,278
4	Claus Plant	3,824
5	H_2S Removal Unit	3,254
6	Heater, Air Compressor and Turbine	7,473
7	Other Offsites	22,446
TPI		82,179
Credit		40,000
TPI Adjusted		42,006

	Coal Price		
	$5/T	$10/T	$15/T
TCR: M$	106,017	114,076	122,135
TCR Adjusted M$	58,817	66,876	74,935
Raw Materials and Utilities Costs Less Credits, M$/YR			
ACOAL: M$	20,148	40,296	60,444
ASULFUR: $5/LT		(566)	
UTILITIES		4,271	
ANR: M$/YR	23,853	44,001	64,149
Total Annual Production Cost			
0.217 TPI		9,115	
2.1 (TO) (CO) (1 + F)		11,157	
1.04 ANR	24,807	45,761	66,715
TAC: M$/YR	45,079	66,033	86,987
: ¢/MM BTU	73.9	108.3	142.7

Figure 7.1

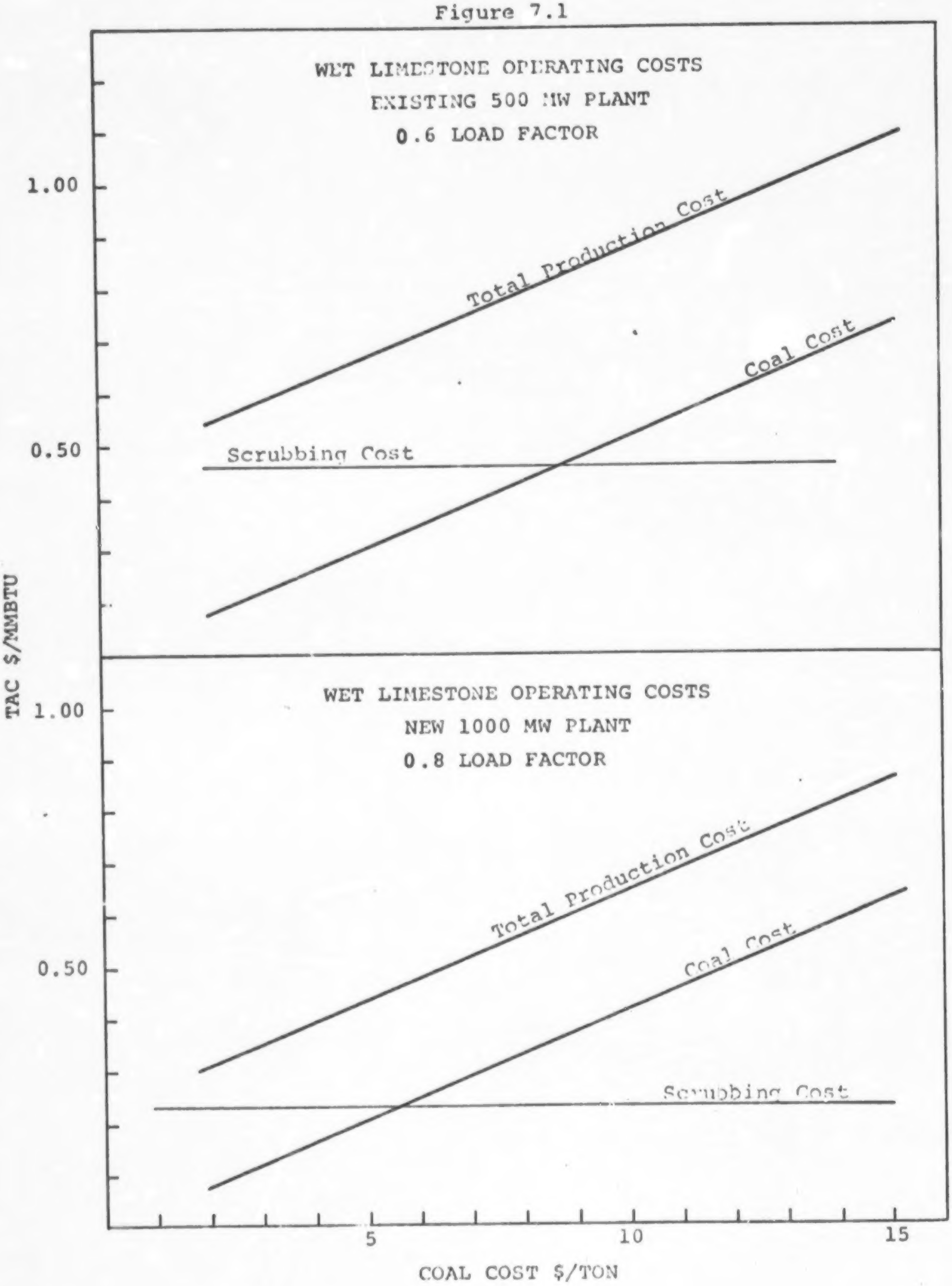

WET LIMESTONE OPERATING COSTS
EXISTING 500 MW PLANT
0.6 LOAD FACTOR
Total Production Cost
Coal Cost
Scrubbing Cost
1.00
0.50
TAC $/MMBTU
WET LIMESTONE OPERATING COSTS
NEW 1000 MW PLANT
0.8 LOAD FACTOR
Total Production Cost
Coal Cost
Scrubbing Cost
1.00
0.50
5
10
15
COAL COST $/TON

Figure 7.2

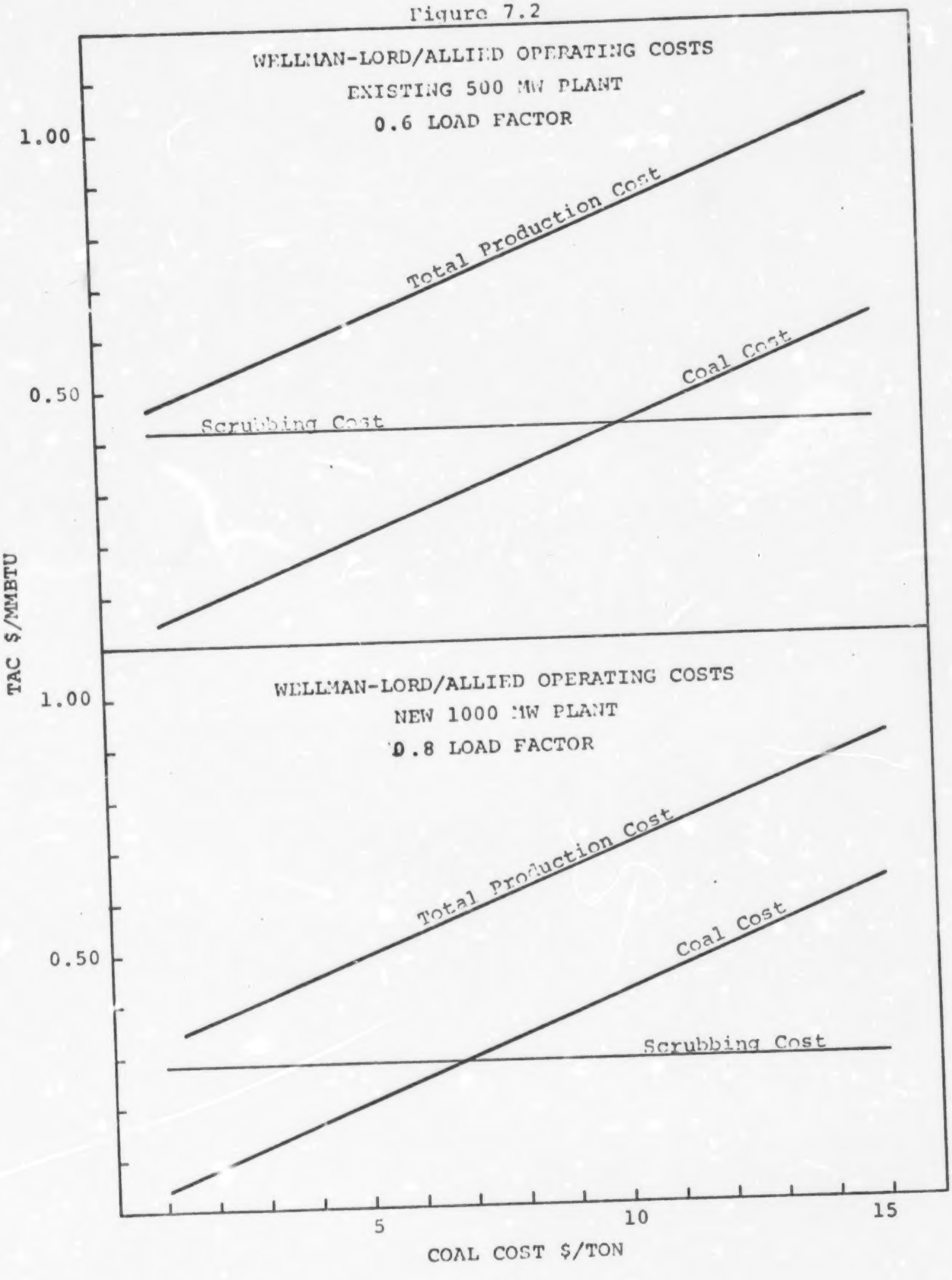
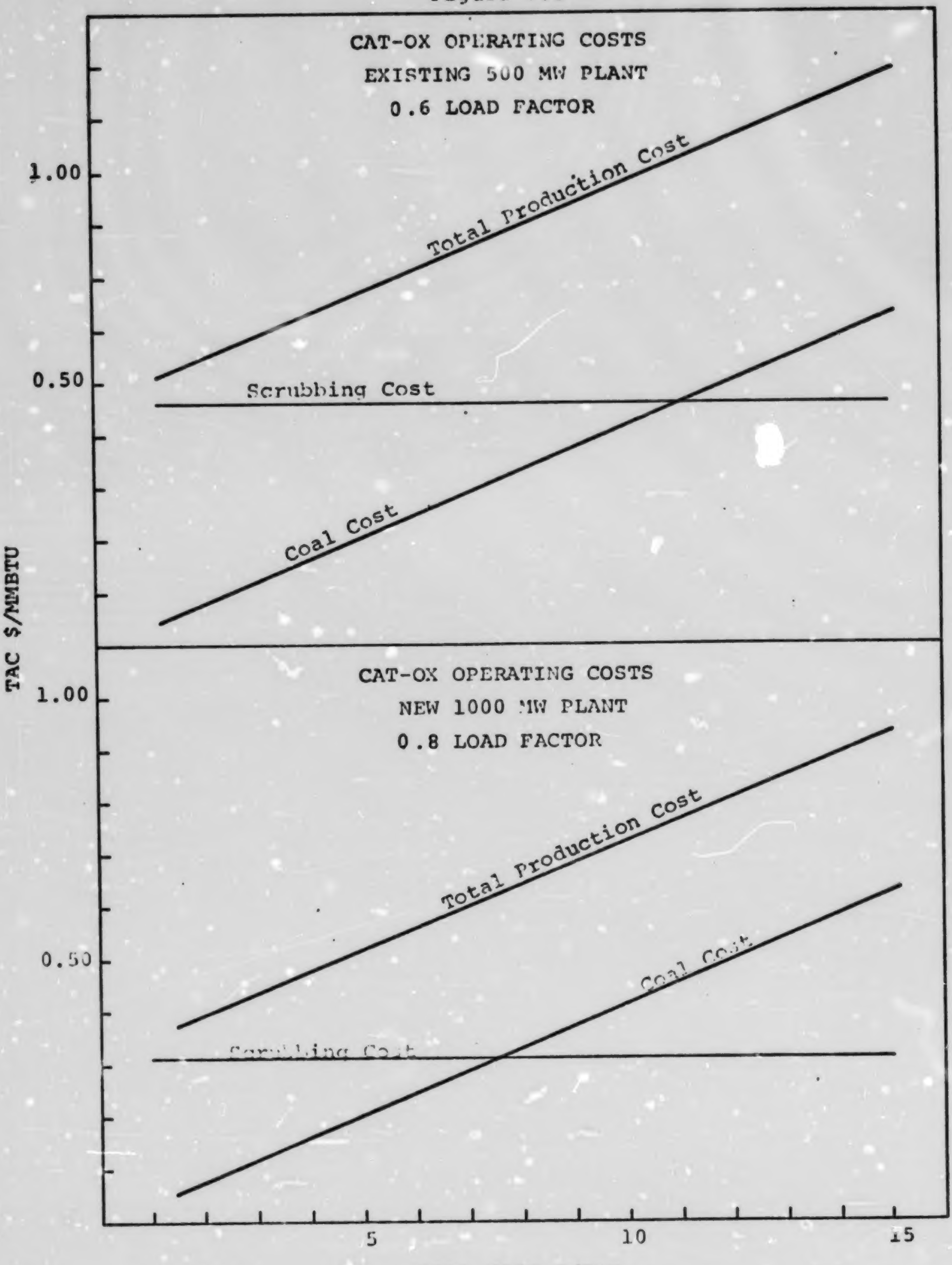

WELLMAN-LORD/ALLIED OPERATING COSTS
EXISTING 500 MW PLANT
0.6 LOAD FACTOR
Total Production Cost
Coal Cost
Scrubbing Cost
1.00
0.50
WELLMAN-LORD/ALLIED OPERATING COSTS
NEW 1000 MW PLANT
0.8 LOAD FACTOR
Total Production Cost
Coal Cost
Scrubbing Cost
1.00
0.50
TAC $/MMBTU
5
10
15
COAL COST $/TON

Figure 7.3



CAT-OX OPERATING COSTS
EXISTING 500 MW PLANT
0.6 LOAD FACTOR
1.00
0.50
Total Production Cost
Scrubbing Cost
Coal Cost
TAC $/MMBTU
CAT-OX OPERATING COSTS
NEW 1000 MW PLANT
0.8 LOAD FACTOR
1.00
0.50
Total Production Cost
Coal Cost
Scrubbing Cost
5
10
15
COAL COST $/TON

Figure 7.4

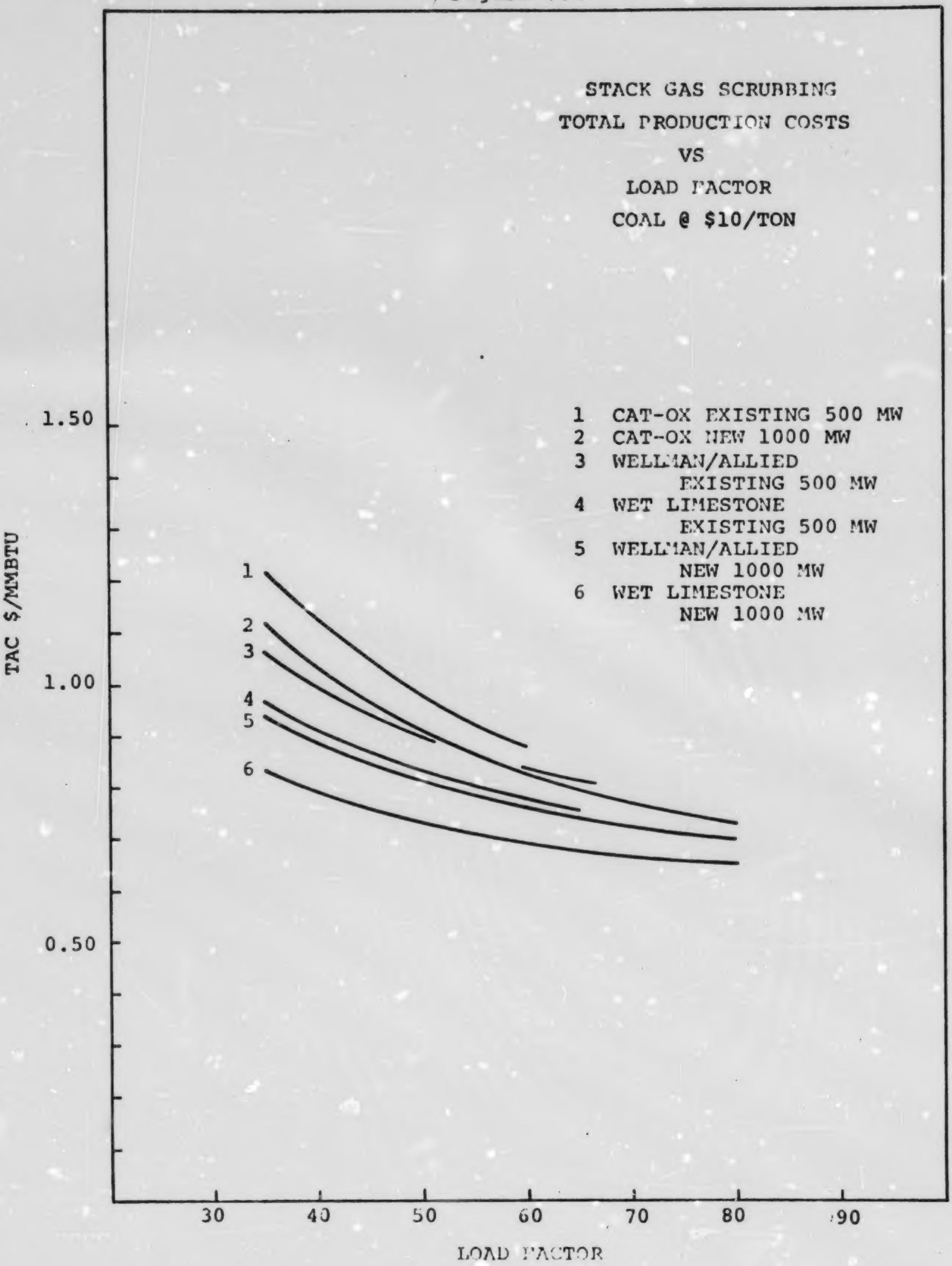

STACK GAS SCRUBBING
TOTAL PRODUCTION COSTS
VS
LOAD FACTOR
COAL @ $10/TON
1 CAT-OX EXISTING 500 MW
2 CAT-OX NEW 1000 MW
3 WELLMAN/ALLIED EXISTING 500 MW
4 WET LIMESTONE EXISTING 500 MW
5 WELLMAN/ALLIED NEW 1000 MW
6 WET LIMESTONE NEW 1000 MW
1
2
3
4
5
6
TAC $/MMBTU
1.50
1.00
0.50
30
40
50
60
70
80
90
LOAD FACTOR

Figure 7.5

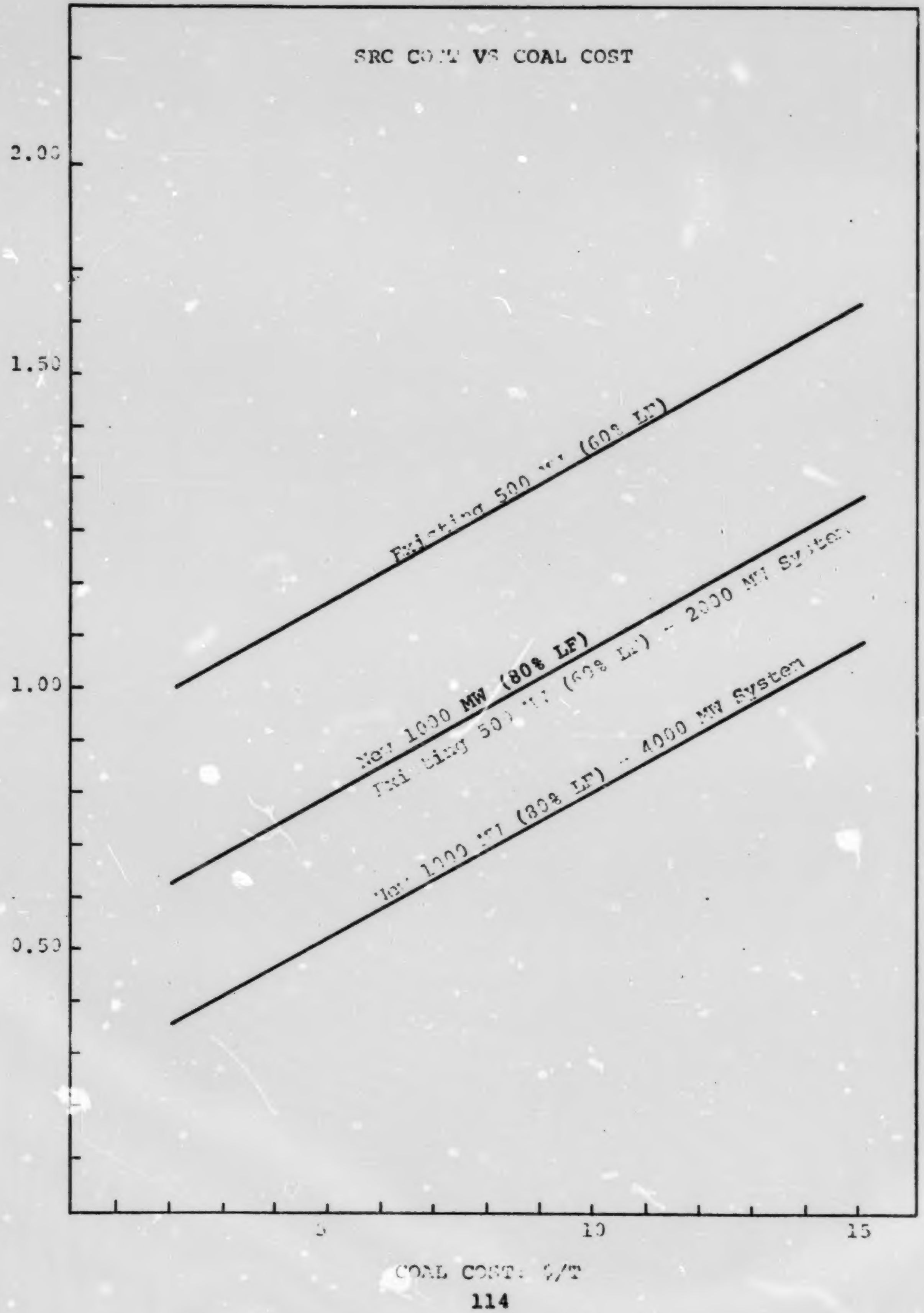
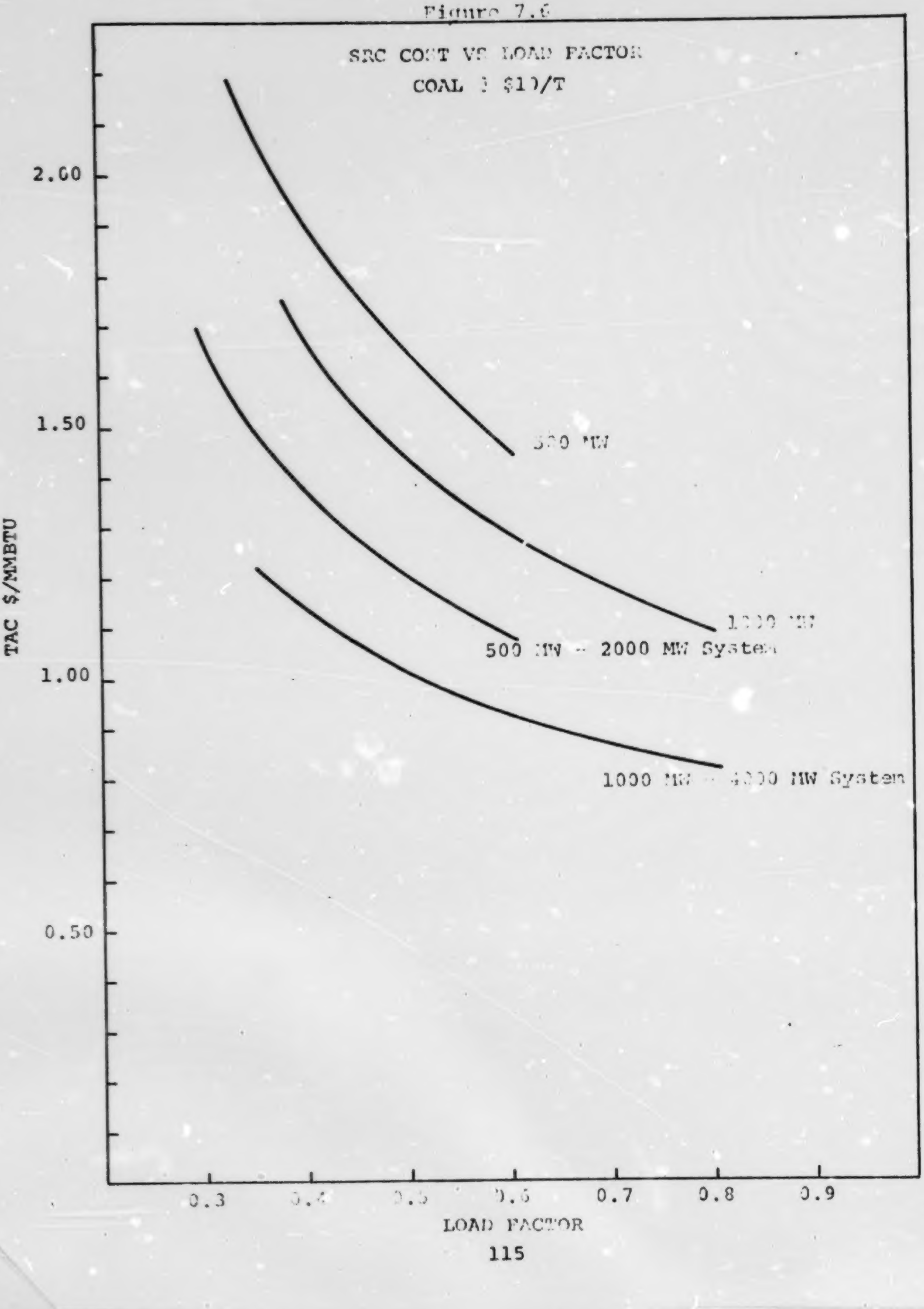

SRC COST VS COAL COST
2.00
1.50
1.00
0.50
Existing 500 MW (60% LF)
New 1000 MW (80% LF)
Existing 500 MW (60% LF) - 2000 MW System
New 1000 MW (80% LF) - 4000 MW System
5
10
15
COAL COST: $/T

Figure 7.6


SRC COST VS LOAD FACTOR
COAL @ $10/T
500 MW
1000 MW
500 MW – 2000 MW System
1000 MW – 4000 MW System
TAC $/MMBTU
2.00
1.50
1.00
0.50
0.3
0.4
0.5
0.6
0.7
0.8
0.9
LOAD FACTOR

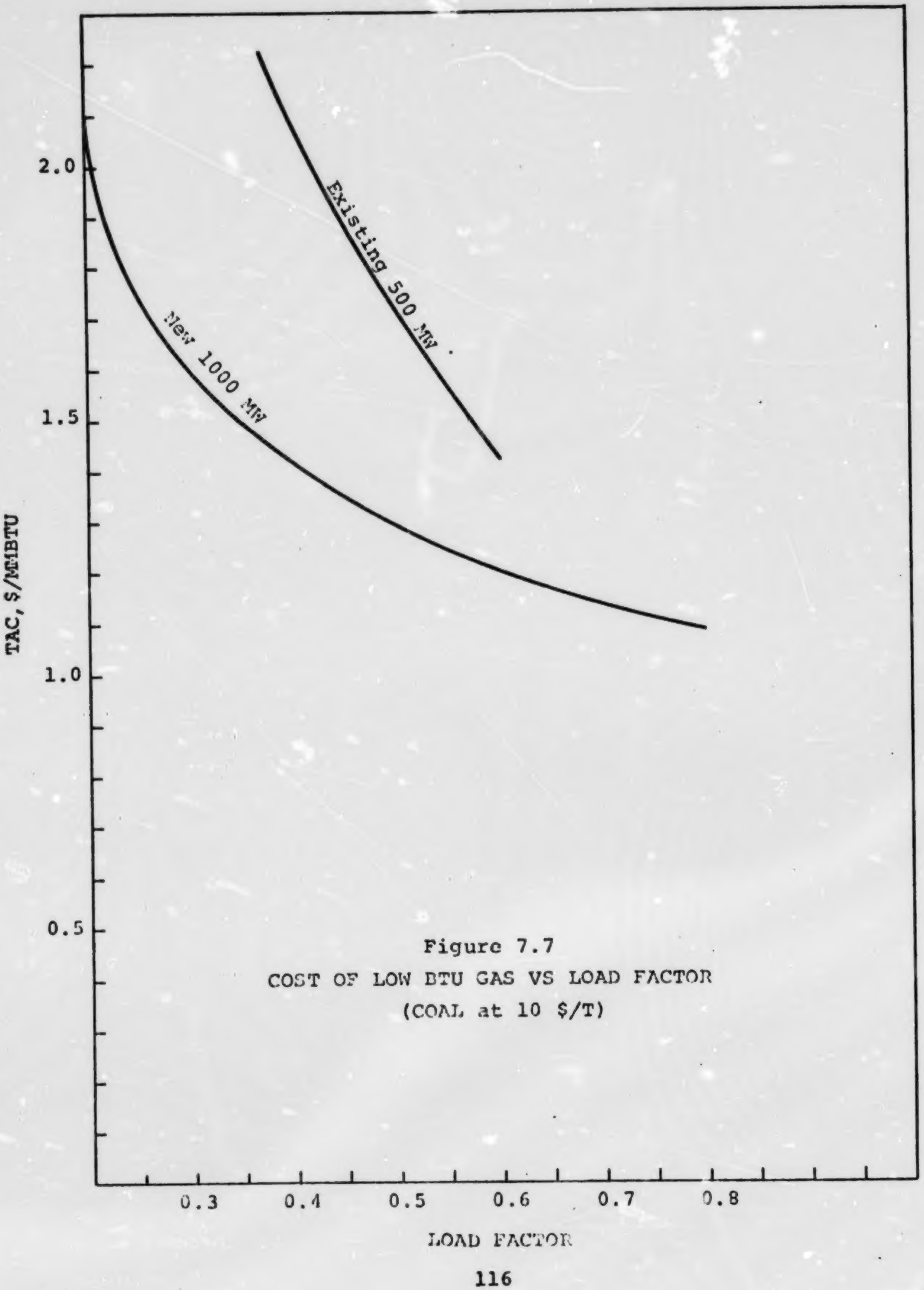


Figure 7.7
COST OF LOW BTU GAS VS LOAD FACTOR
(COAL at 10 $/T)

Figure 7.8

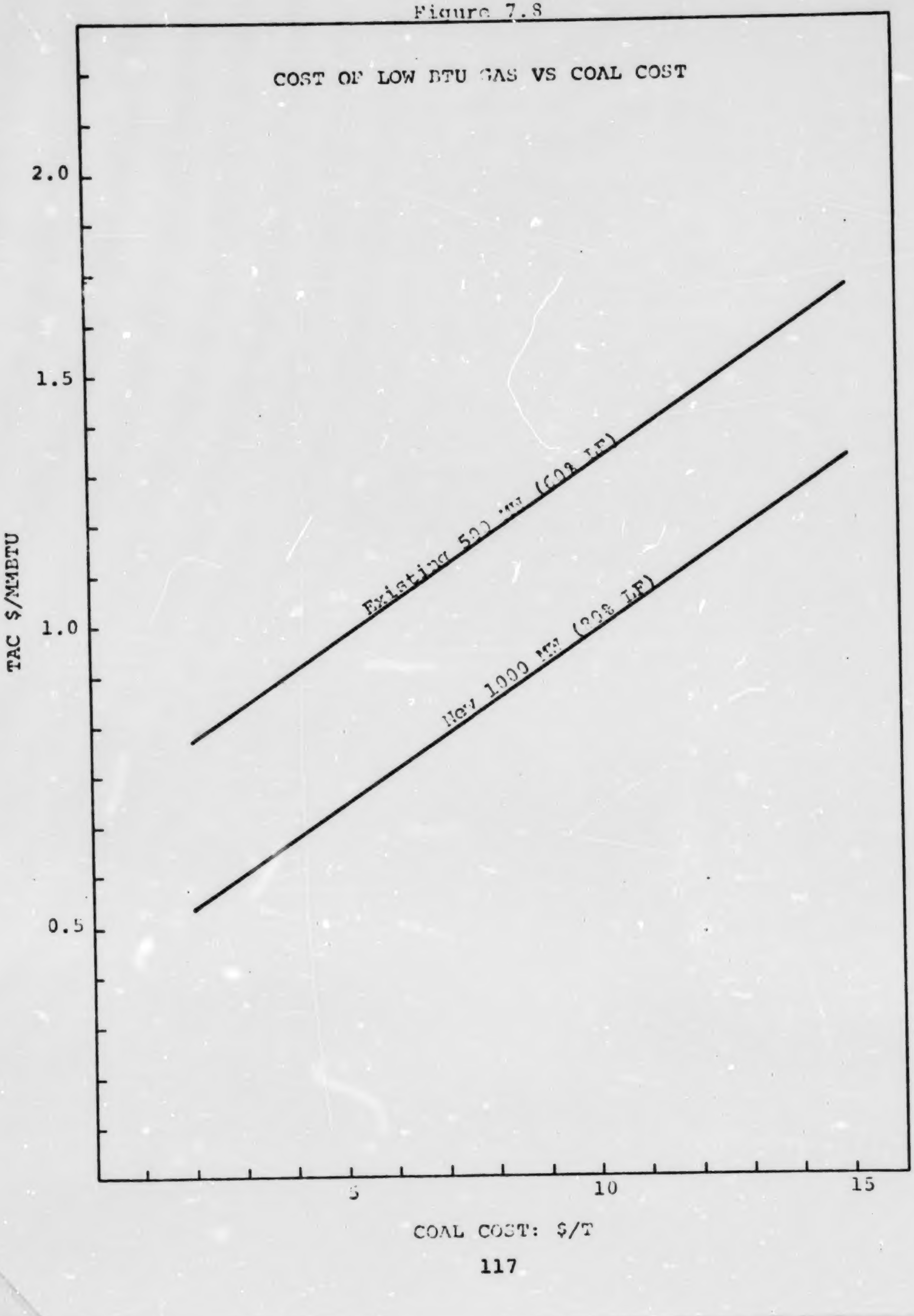

COST OF LOW BTU GAS VS COAL COST
TAC $/MMBTU
2.0
1.5
1.0
0.5
Existing 500 MW (60% LF)
New 1000 MW (80% LF)
5
10
15
COAL COST: $/T

Figure 7.9

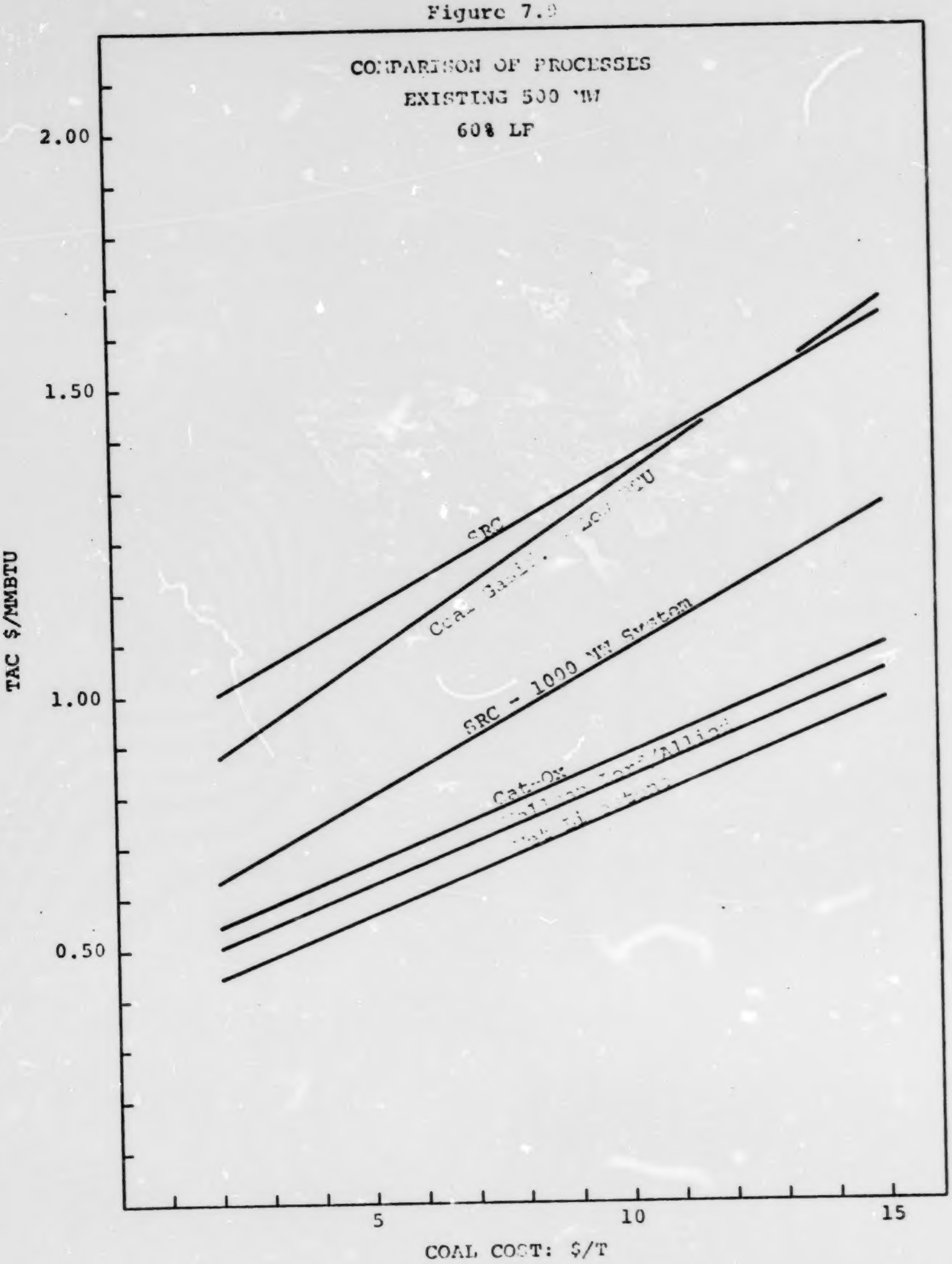
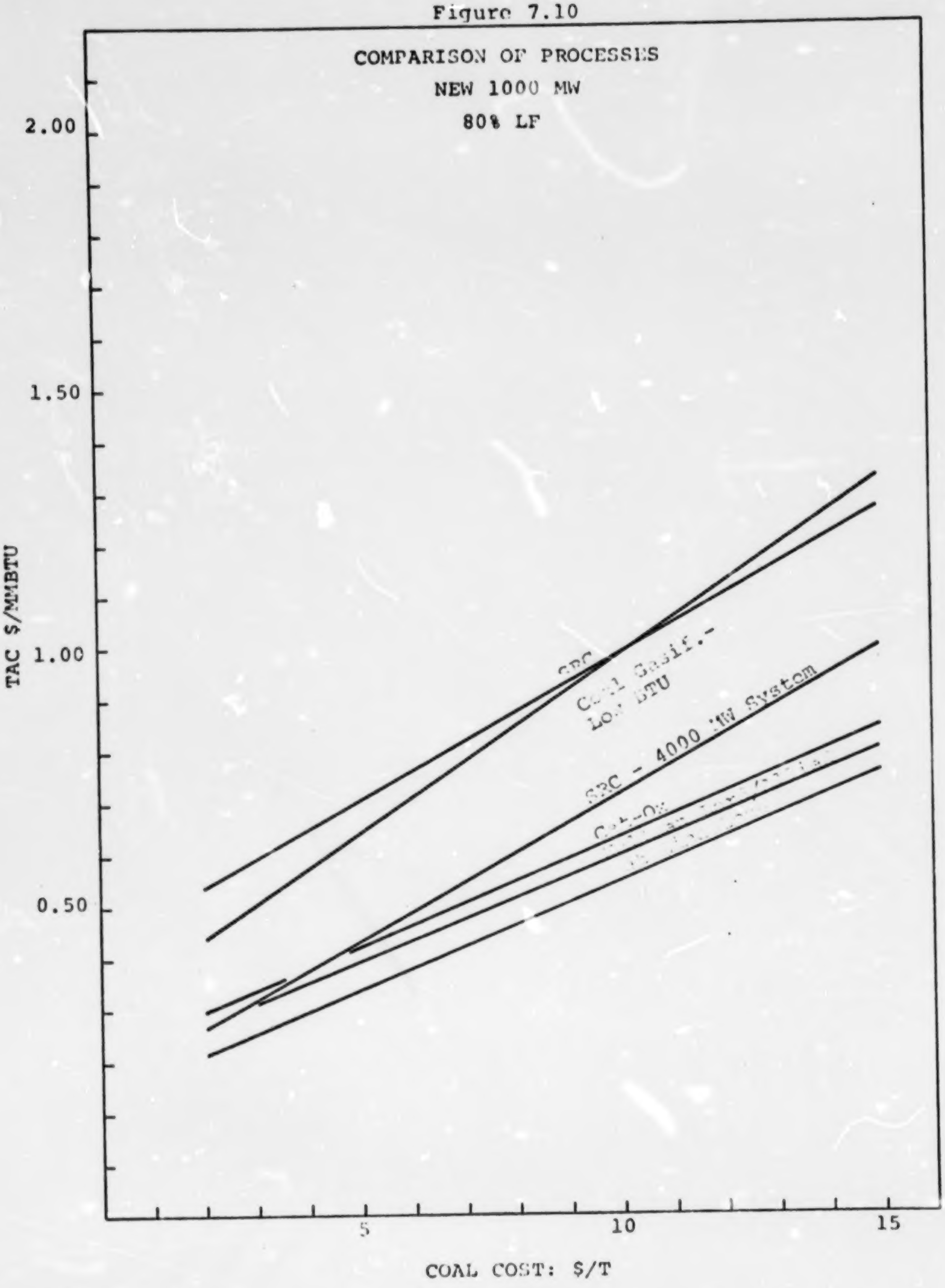

COMPARISON OF PROCESSES
EXISTING 500 MW
60% LF
TAC $/MMBTU
2.00
1.50
1.00
0.50
SRC
Coal Gasif. - Low BTU
SRC - 1000 MW System
Cat-Ox
Wellman Lord/Allied
Wet Lime System
5
10
15
COAL COST: $/T

Figure 7.10


COMPARISON OF PROCESSES
NEW 1000 MW
80% LF
2.00
1.50
1.00
0.50
TAC $/MMBTU
SRC
Coal Gasif.-
Low BTU
SRC - 4000 MW System
Cat-Ox
5
10
15
COAL COST: $/T

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9. APPENDIX

(Excerpted from reference 3)

4. THE GENERAL MODEL

4.1 The General Process Model

The plants in the models have, as far as possible, been made self-contained apart from the intake of basic raw feed materials; i.e., the plant should not be buying natural gas or electricity. If possible, it should not even be buying desulfurized fuel oil since supply cannot be assumed. There are obviously exceptions if the plant is an addition to a larger conventional plant; e.g., with stack gas scrubbing for a power plant it would be illogical not to assume a supply of power. In general, a large plant having a coal feed will generate its own power, steam and heat requirements by burning coal and scrubbing the stack gases.

It was not a primary concern to provide special chemical by-products from any process, but to avoid additional treatment facilities for impure materials by routing these side streams back to the plant fuel supply where possible. This approach simplifies the models and minimizes the effect of credits for special chemical by-products on the plant costs.

The cost of equipment and raw material, utility and waste product quantities have all been related to one or more basic process parameters; e.g., in the stack gas scrubbing models, the basic process parameters are flue gas flow rate and sulfur content of the fuel. For a plant producing high quality fuel, the basic process parameters are product flow rate and properties of the raw feed materials.

Where possible, equipment costs were related directly to the basic process parameters. However, the format of some of the estimates used to develop the models prevented this. In these cases, the available cost information was carefully examined relative to the General Cost Model to determine exactly what the costs included.

The equipment costs were extracted from these estimates by using the relationships between construction labor costs, other material costs and equipment costs given in the General Cost Model.

Each plant design was examined to fix maximum train sizes for each group of equipment. It has been assumed that N trains cost N times the cost of one train. Where a plant is largely made up of several trains, size variations were only taken in increments of their size.

For the smaller plants, it was possible to examine the cost of every item of equipment and assign an exponent of size to give cost variations. However, for the larger plants, whole sections have been grouped together. The following is given as a general guide to the exponents for equipment cost vs. size (9,14,21):

	n, $\left[\frac{\text{Cost}_2}{\text{Cost}_1} = \left(\frac{\text{Size}_2}{\text{Size}_1}\right)^n\right]$
Increasing number of trains of equipment	1.0
Blowers	0.9
Solids grinding equipment	0.8
Steam generation equipment	0.8
Process furnaces and reformers	0.7
Compressors	0.7
Power generation equipment	0.7
Solids handling equipment	0.6
Offsites	0.6
Other process units	0.6

4.2 The General Cost Model

4.2.1 Bases For Costs

All costs in the models are those in existence at the end of 1973. To update prior cost information used in the construction of the models, an annual inflation multiplication factor of 1.05 has been used. All costs other than unit costs for labor, raw materials, etc., are shown in thousands of dollars (M$).

The direct field construction labor cost, L, and the direct cost of operating labor, CO, both refer to a Gulf Coast (Houston) location. For any other location, they are adjusted through the use of a location factor, F, which is explained in section 4.3.

Whenever possible in the development of the cost models discussed in this report, major equipment costs, E, have been related to plant size variations. The reference values of E have been taken from actual plant cost estimates when these were available. Sometimes, however, the cost estimates were not available in such a detailed breakdown. In such cases, the relationships developed in the General Cost Model were used to analyze the cost data. The relationships in the General Cost Model were developed based on procedures reported and recommended in the literature (9,13) and on Kellogg's general experience.

4.2.2 Capital Cost Model

Major equipment costs, E, represent the cost of major equipment delivered to the site, but not located, tied-in to piping, instruments, etc., or commissioned. It includes material costs only. Major equipment is defined to include furnaces, heat exchangers, converters, reactors, towers, drums and tanks, pumps, compressors, transportation and conveying equipment, special equipment (filters, centrifuges, dryers, agitators, grinding equipment, cyclones, etc.), and major gas ductwork.

Other material costs, M, represent the cost of piping, electrical, process instrumentation, paint, insulation, foundations, concrete structures, and structural steel for equipment support. It does not include such items as site preparation, steel frame structures, process buildings, cafeterias, control rooms, shops, offices, etc.

M has been taken as a fixed fraction of E. Whenever possible, this fraction has been determined from an estimate covering the particular plant under consideration. This fraction is often different for each section of the plant. If particular details were not available, the following relationships have been assumed (9):

Solids handling plant:	M = 0.40E
Chemical process plant:	M = 0.80E

Direct field construction labor costs, L, are based on Gulf Coast rates and productivities. Again, L has been taken as a fixed fraction of E. Whenever possible, it has been derived from an estimate covering the particular plant under consideration. This fraction is often different for each section of the plant. If particular details were not available, the following relationships have been assumed (9):

Solids handling plant:	L = 0.40E
Chemical process plant:	L = 0.60E

Indirect costs associated with field labor have been assumed as follows:

Fringe benefits and payroll burden	= 0.12 L
Field administration, supervision temporary facilities	= 0.17 L
Construction equipment and tools	= 0.14 L
Total field labor indirect costs	= 0.43 L

Home office engineering includes home office construction, engineering and design, procurement, client services, accounting, cost engineering, travel and living expenses, reproduction and communication. This could range from under 10% to almost 20% of the major equipment and other material costs. In the model, this has been assumed to be 15% of the total direct material cost (E + M).

The bare cost of the plant, BARC, is defined as the sum of equipment costs, other material costs, construction labor and labor indirects, and home office engineering. For a Gulf Coast location, it is given by:

$$\begin{aligned} BARC &= E + M + L + 0.43\ L + 0.15\ (E + M) \\ &= 1.15\ (E + M) + 1.43\ L \end{aligned}$$

For any other location, it is given by:

$$BARC = 1.15\ (E + M) + 1.43\ L \cdot F$$

where F is the location factor (see section 4.3).

Taxes and insurance can be 1-4% of the bare cost. In the model, they have been assumed to be 2%. Contractor's overheads and profit could depend on several factors, but are generally in the range of 6-13% of the bare cost. A value of 10% was chosen for the model.

A contingency has been included in the model and is expressed as a fraction of the bare cost. It represents the degree of uncertainty in the process design and the cost estimate. The contingency, CONTIN, could range from zero for a well-established process to 0.20 or more for a process still under development.

The total plant investment, TPI, is defined as the sum of the bare cost (including contingency), taxes and insurance, and contractor's overheads and profit. It is therefore given by:

$$\begin{aligned} TPI &= (1.0 + CONTIN)\ BARC + 0.02\ (1.0 + CONTIN)\ BARC \\ &\quad + 0.10\ (1.0 + CONTIN)\ BARC \\ &= 1.12\ (1.0 + CONTIN)\ BARC \end{aligned}$$

In order to obtain the total capital required for construction of a particular plant, some additional costs should be added to the total plant investment. These costs are:

1. Start-up costs
2. Working capital
3. Interest during construction

Start-up costs, STC, have been assumed to be 20% of the total net annual operating cost, AOC (see section 4.2.3 for explanation of AOC). Thus:

$$STC = 0.20\ AOC$$

Working capital, WKC, is required for raw materials inventory, plant materials and supplies, etc. For simplification, it has also been assumed to be 20% of the total net annual operating cost, AOC.

Thus:

$$WKC = 0.20\ AOC$$

Interest during construction, IDC, obviously increases with the length of the construction period which, to some extent, is a function of the size of the plant. The construction of plants the size of the stack gas scrubbing units is now taking about 2-3 years and projects of the magnitude and complexity of a substitute natural gas plant or a power station are taking 4-5 years. Two different values for the interest during construction have therefore been assumed. The first is intended to be used for stack gas scrubbing units fitted to existing power plants or for constructions well under $100 million:

$$\mathrm{IDC} = 0.12\ \mathrm{TPI}^{*}$$

The second is for the larger, more complex plants such as substitute natural gas, solvent refined coal, and power plants:

$$\mathrm{IDC} = 0.18\ \mathrm{TPI}^{*}$$

The total capital required, TCR, is equal to the sum of the total plant investment, start-up costs, working capital, and interest during construction.

Thus:

$$\mathrm{TCR} = \mathrm{TPI} + \mathrm{STC} + \mathrm{WKC} + \mathrm{IDC}$$

For stack gas scrubbing units, this can be reduced to:

$$\begin{aligned}\mathrm{TCR} &= \mathrm{TPI} + 0.20\ \mathrm{AOC} + 0.20\ \mathrm{AOC} + 0.12\ \mathrm{TPI}\\ &= 1.12\ \mathrm{TPI} + 0.40\ \mathrm{AOC}\end{aligned}$$

For the larger plants, this can be reduced to:

$$\begin{aligned}\mathrm{TCR} &= \mathrm{TPI} + 0.20\ \mathrm{AOC} + 0.20\ \mathrm{AOC} + 0.18\ \mathrm{TPI}\\ &= 1.18\ \mathrm{TPI} + 0.40\ \mathrm{AOC}\end{aligned}$$

From section 4.2.3, AOC is calculated from:

$$\mathrm{AOC} = 0.078\ \mathrm{TPI} + 2.0\ \mathrm{TO}\cdot\mathrm{CO}\ (1.0 + F) + \mathrm{ANR}$$

---

*See Appendix A for derivation of equation where TO = total number of shift operators

ANR = Annual cost of raw materials, utilities, and waste disposal, less by-product credits.

Therefore, for stack gas scrubbing units, the equation for the total capital required becomes:

$$\begin{aligned} TCR &= 1.12\ TPI + 0.40\ [0.078\ TPI + 2.0\ TO \cdot CO\ (1.0 + F) + ANR] \\ &= 1.12\ TPI + 0.03\ TPI + 0.8\ TO \cdot CO\ (1.0 + F) + 0.40\ ANR \\ &= 1.15\ TPI + 0.8\ TO \cdot CO\ (1.0 + F) + 0.40\ ANR \end{aligned}$$

For the larger plants, the equation for the total capital required becomes:

$$\begin{aligned} TCR &= 1.18\ TPI + 0.40\ [0.078 TPI + 2.0\ TO \cdot CO\ (1.0 + F) + ANR] \\ &= 1.18\ TPI + 0.03\ TPI + 0.8\ TO \cdot CO\ (1.0 + F) + 0.4\ ANR \\ &= 1.21\ TPI + 0.8\ TO \cdot CO\ (1.0 + F) + 0.4\ ANR \end{aligned}$$

The buildup of costs to determine the total capital required is illustrated in Figure 4.1.

4.2.3 Operating Cost Model

The total net annual operating cost, AOC, is the total cost of operating the plant less the credits from the sale of by-products. It does not include return of capital, payment of interest on capital, income tax on equity returns or depreciation. The total net annual operating cost is made up of the following items:

1. Annual cost of raw materials, utilities, and waste disposal, less by-product credits
2. Annual cost of operating labor and supervision
3. Annual cost of maintenance labor and supervision
4. Annual cost of plant supplies and replacements
5. Annual cost of administration and overheads
6. Annual cost of local taxes and insurance

The annual cost of raw materials, utilities, and waste disposal, less by-product credits, ANR, is clearly a function of the particular process under consideration. It is given by different relationships for each model.

The total number of operators employed on all shifts, TO, is different for each process and is either given as an equation or number for each particular model. It has been assumed that each operator works 40 hours per week for 50 weeks per year (2000 hours per year). If CO is the hourly rate for an operator (Gulf Coast basis), then the annual cost of operating labor is given by:

$$\text{Operating labor (Gulf Coast)} = \frac{TO \cdot 2000 \cdot CO}{1000}$$

$$= 2\ TO \cdot CO \qquad \text{M\$/yr}$$

The annual cost of operating labor for any other location has been assumed to be:

$$\text{Operating labor} = 2\ TO \cdot CO\ (0.5 + 0.5\ F)$$

Supervision was assumed to be 15% of operating labor. Thus, the total cost of operating labor and supervision, AOL, is given by:

$$\begin{aligned} AOL &= 1.15\ [2\ TO \cdot CO\ (0.5 + 0.5\ F)] \\ &= 2.3\ TO \cdot CO\ (0.5 + 0.5\ F) \end{aligned}$$

The annual cost of maintenance labor has been assumed to be 1.5% of the total plant investment. Maintenance supervision is 15% of maintenance labor. Therefore, the total annual cost of maintenance labor and supervision, AML, is:

$$AML = 1.15\ (0.015\ TPI)$$
$$= 0.018\ TPI\ \text{(rounded up)}$$

Plant supplies and replacements include charts, cleaning supplies, miscellaneous chemicals, lubricants, paint, and replacement parts such as gaskets, seals, valves, insulation, welding materials, packing, balls (grinding), vessel lining materials, etc. The annual cost of plant supplies and replacements, APS, has been assumed to be 2% of the total plant investment. Thus:

$$APS = 0.02\ TPI$$

Administration and overheads include salaries and wages for administrators, secretaries, typists, etc., office supplies and equipment, medical and safety services, transportation and communications, lighting, janitorial services, plant protection, payroll overheads, employee benefits, etc. The annual cost of administration and overheads, AOH, has been assumed to be 70% of the annual operator, maintenance labor, and total supervision costs. Thus:

$$AOH = 0.70\ [2.3\ TO \cdot CO\ (0.5 + 0.5F) + 0.018\ TPI]$$
$$= 1.7\ TO \cdot CO\ (0.5 + 0.5F) + 0.013\ TPI\ \text{(rounded up)}$$

Local taxes and insurance include property taxes, fire and liability insurance, special hazards insurance, business interruption insurance, etc. The annual local taxes and insurance, ATI, have been assumed to be 2.7% of the total plant investment. Thus:

$$ATI = 0.027\ TPI$$

The total net annual operating cost, AOC, is therefore given by:

$$\begin{aligned} AOC &= ANR + AOL + AML + APS + AOH + ATI \\ &= ANR + 2.3\ TO \cdot CO\ (0.5 + 0.5F) + 0.018\ TPI \\ &\quad + 0.02\ TPI + 1.7\ TO \cdot CO\ (0.5 + 0.5F) + 0.013\ TPI \\ &\quad + 0.027\ TPI \\ &= 0.078\ TPI + 4.0\ TO \cdot CO\ (0.5 + 0.5F) + ANR \\ &= 0.078\ TPI + 2.0\ TO \cdot CO\ (1.0 + F) + ANR \end{aligned}$$

In order to obtain the total annual production cost, the following items must be added to the total net annual operating cost:

1. depreciation
2. average yearly interest on borrowed capital
3. average yearly net return on equity
4. average yearly income tax

The straight-line method was used to determine depreciation, based on the total capital required less the working capital. For stack gas scrubbing units (15 year life), the annual depreciation, ACR, is:

$$\begin{aligned} ACR &= 1/15\ (TCR - WKC) \\ &= 0.067\ (TCR - 0.20\ AOC) \end{aligned}$$

For substitute natural gas and solvent refined coal plants (20 year life), it is given by:

$$ACR = 0.050\ (TCR - 0.20\ AOC)$$

For power plants, both conventional and combined cycle (28 year life), it is:

$$ACR = 0.036\ (TCR - 0.20\ AOC)$$

Interest on debt and return on equity are calculated following a procedure recommended in the literature (13) and illustrated in Appendix A. The procedure assumes a fixed debt-to-equity ratio, an interest rate on debt, and the required net (after tax) rate of return on equity. Interest on debt and return on equity are calculated over the plant life, and the yearly average is expressed as a percentage of the total capital required (TCR). Assuming a 75%/25% debt-to-equity ratio, a 9% per year interest rate, and a 15% per year net rate of return on equity, the annual interest and return, AIC, is given by:

$$AIC = 0.054\ TCR$$

Federal income tax is the average yearly income tax over the plant life, expressed as a percentage of the total capital required. The calculation of income tax is illustrated in Appendix A. Based on the assumptions listed in the preceding paragraph and an assumed tax rate of 48%, the annual federal income tax, AFT, is given by :

$$AFT = 0.018\ TCR$$

The total annual production cost, TAC, is given by:

$$TAC = AOC + ACR + AIC + AFT$$

For stack gas scrubbing plants, this can be reduced as follows:

$$\begin{aligned} TAC &= AOC + 0.067\ (TCR - 0.20\ AOC) + 0.054\ TCR + 0.018\ TCR \\ &= AOC + 0.067\ TCR - .013\ AOC + 0.054\ TCR + 0.018\ TCR \\ &= 0.139\ TCR + 0.99\ AOC \end{aligned}$$

Substituting for TCR and AOC from preceeding equations:

$$\begin{aligned} TAC &= 0.139\ [1.15\ TPI + 0.8\ TO \cdot CO\ (1.0 + F) + 0.40\ ANR] \\ &\quad + 0.99\ [0.078\ TPI + 2.0\ TO \cdot CO\ (1.0 + F) + ANR] \\ &= 0.237\ TPI + 2.1\ TO \cdot CO\ (1.0 + F) + 1.04\ ANR \end{aligned}$$

Making the appropriate substitutions, the total annual production cost for substitute natural gas and solvent refined coal plants is:

$$TAC = 0.225\ TPI + 2.1\ TO \cdot CO\ (1.0 + F) + 1.04\ ANR$$

For power plants, this equation becomes:

$$TAC = 0.208\ TPI + 2.1\ TO \cdot CO\ (1.0 + F) + 1.04\ ANR$$

The buildup of costs to determine the total annual production cost is illustrated in Figure 4.2.

4.3 Effect of Location on Plant Cost

The cost models have been developed using U.S. Gulf Coast 1973 costs as a basis. In order to predict plant costs for other locations, factors have been developed which relate construction labor costs at various locations to Gulf Coast labor costs. By multiplying the field labor construction portion of plant cost by this location factor, the total plant cost is adjusted to the desired location.

Labor rates for different crafts were obtained from the literature (10) and escalated to the end of 1973. Using an average craft mix obtained from in-house information (12), an average construction labor rate was obtained for each location. Productivity factors for the various locations, also obtained from in-house data, were used to create the rate for equal work output. These rates were then normalized, using Houston (Gulf Coast) as a basis, to yield relative field labor construction costs.

Table 4.1 lists the relative labor costs determined for twenty cities. They range from 1.0 for Houston to 2.08 for New York. Costs are generally highest in the Northeastern quarter of the country and lowest in the South. These factors are shown on a map of the U.S. in Figure 4.3.

Table 4.2 lists average location factors for each state. Allowance has been made in the factor for the importation of temporary labor to the more remote states. The factors are shown on a map of the U.S. in Figure 4.4.

Figure 4.5 gives the relationship between major equipment cost, E, total plant investment, TPI, and location factor, F, when the contingency, CONTIN, is zero.

4.4 Nomenclature

E	Major equipment costs	M$
M	Other material costs	M$
L	Direct field labor costs (Gulf Coast)	M$
BARC	Bare cost	M$
F	Location factor	
CONTIN	Contingency	
TPI	Total plant investment	M$
STC	Start-up costs	M$
WKC	Working capital	M$
IDC	Interest during construction	M$
TCR	Total capital required	M$
ANR	Annual cost of raw materials, utilities, and waste disposal, less by-product credits	M$/year
AOL	Annual cost of operating labor and supervision	M$/year
AML	Annual cost of maintenance labor and supervision	M$/year
APS	Annual cost of plant supplies and replacements	M$/year

AOH	Annual cost of administration and overheads	M$/year
ATI	Annual cost of local taxes and insurance	M$/year
AOC	Total net annual operating cost	M$/year
TO	Total number of shift operators	
CO	Hourly rate for shift operators (Gulf Coast)	$/hour
ACR	Annual depreciation	M$/year
AIC	Annual interest on debt and return on capital	M$/year
AFT	Annual federal income taxes	M$/year
TAC	Total annual production cost	M$/year
COHP	Contractor overhead & profits	M$/year
TAXI	Taxes and insurance	M$/year
FLIC	Field Labor Indirect Cost	M$/year
ENGR	Engineering Fees	M$/year

TABLE 4.1

LOCATION FACTORS FOR MAJOR U.S. CITIES

Location	Location Factor F
Atlanta	1.10
Baltimore	1.41
Birmingham	1.16
Boston	1.23
Chicago	1.52
Cincinnati	1.53
Cleveland	1.86
Dallas	1.07
Denver	1.03
Detroit	1.73
Kansas City	1.37
Los Angeles	1.44
Minneapolis	1.54
New Orleans	1.16
New York	2.08
Philadelphia	1.82
Pittsburgh	1.52
St. Louis	2.01
San Francisco	1.45
Seattle	1.21
Houston	1.00

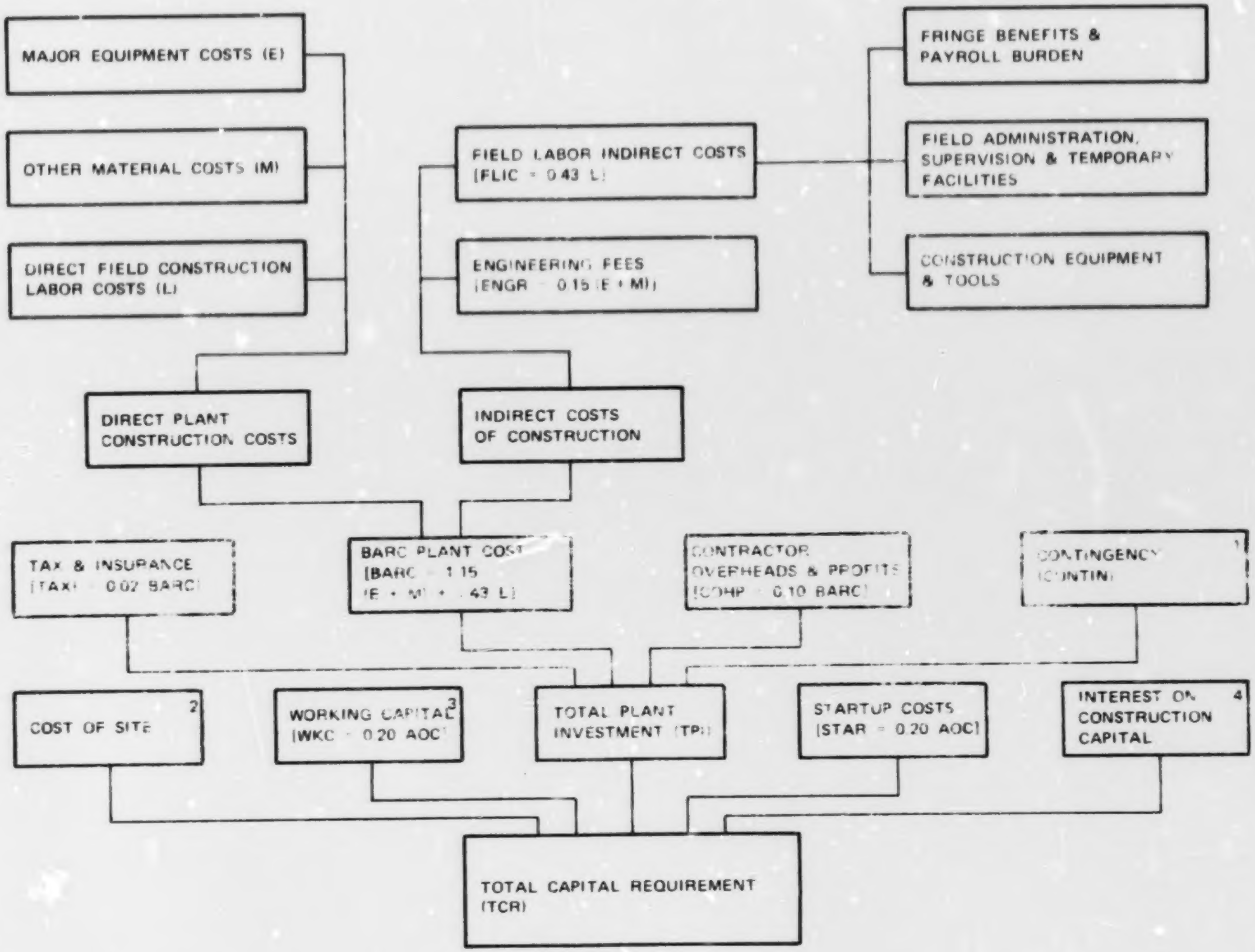
TABLE 4.2

AVERAGE LOCATION FACTORS FOR EACH STATE

State	Location Factor
Alabama	1.2
Alaska	2.1
Arizona	1.3
Arkansas	1.2
California	1.5
Colorado	1.2
Connecticut	1.7
Delaware	1.4
D.C.	1.4
Florida	1.2
Georgia	1.1
Hawaii	2.0
Idaho	1.3
Illinois	1.7
Indiana	1.6
Iowa	1.5
Kansas	1.4
Kentucky	1.5
Louisiana	1.1
Maine	1.2
Maryland	1.4
Massachusetts	1.3
Michigan	1.7
Minnesota	1.5
Mississippi	1.1
Missouri	1.6
Montana	1.3
Nebraska	1.4
Nevada	1.4
New Hampshire	1.2
New Jersey	2.1
New Mexico	1.3
New York	2.1
N. Carolina	1.2
North Dakota	1.3
Ohio	1.6
Oklahoma	1.4
Oregon	1.2
Pennsylvania	1.6
Rhode Island	1.3
S. Carolina	1.1
South Dakota	1.3
Tennessee	1.2
Texas	1.1
Utah	1.2
Vermont	1.2
Virginia	1.4
Washington	1.2
W. Virginia	1.5
Wisconsin	1.5
Wyoming	1.3

FIGURE 4.1

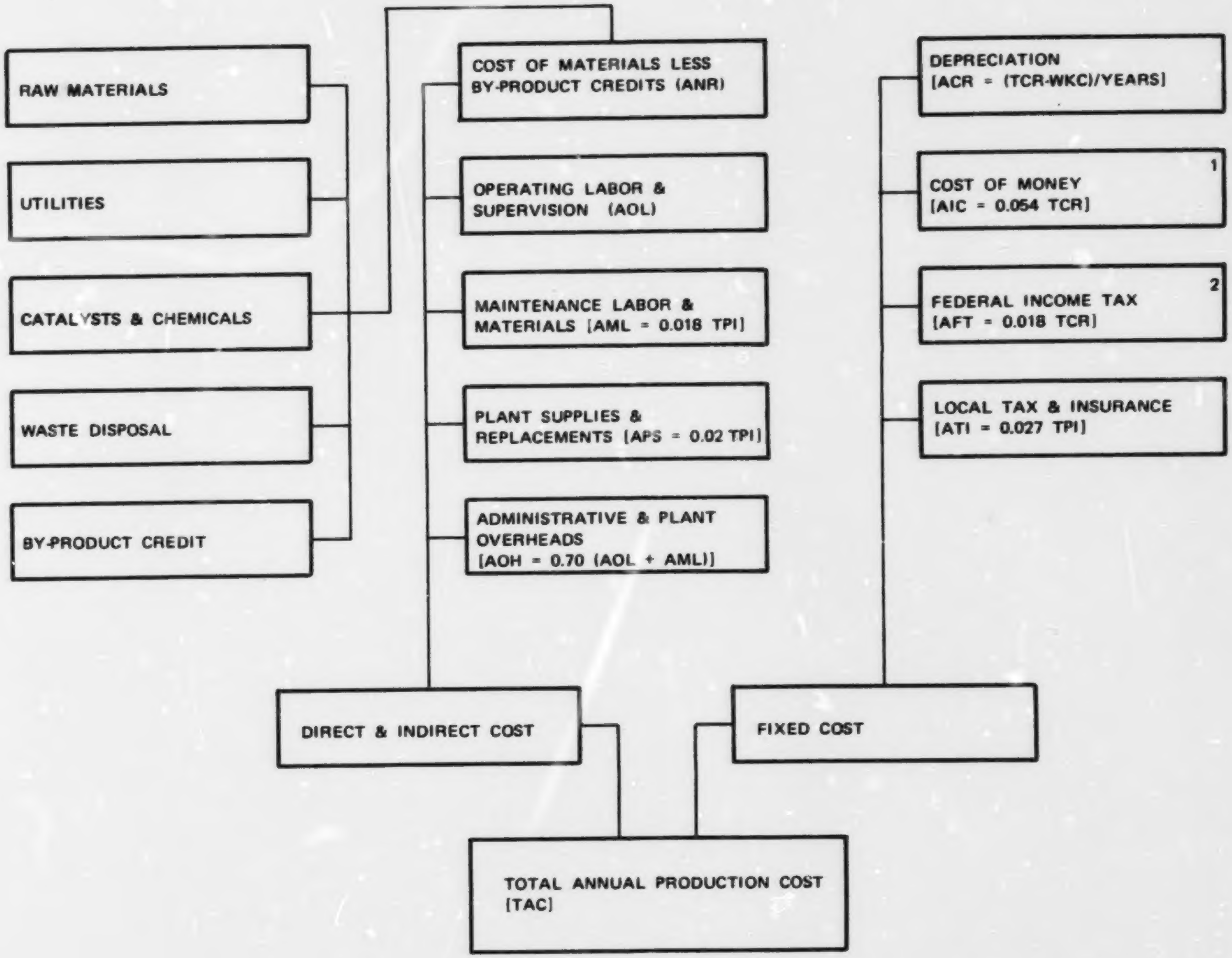
RELATIONSHIP BETWEEN CAPITAL COST FACTORS IN THE GENERAL COST MODEL


MAJOR EQUIPMENT COSTS (E)
OTHER MATERIAL COSTS (M)
DIRECT FIELD CONSTRUCTION LABOR COSTS (L)
DIRECT PLANT CONSTRUCTION COSTS
FIELD LABOR INDIRECT COSTS [FLIC = 0.43 L]
ENGINEERING FEES [ENGR = 0.15 (E + M)]
FRINGE BENEFITS & PAYROLL BURDEN
FIELD ADMINISTRATION, SUPERVISION & TEMPORARY FACILITIES
CONSTRUCTION EQUIPMENT & TOOLS
INDIRECT COSTS OF CONSTRUCTION
TAX & INSURANCE [TAXI = 0.02 BARC]
BARC PLANT COST [BARC = 1.15 (E + M) + .43 L]
CONTRACTOR OVERHEADS & PROFITS [COHP = 0.10 BARC]
CONTINGENCY 1 (CONTIN)
COST OF SITE 2
WORKING CAPITAL 3 [WKC = 0.20 AOC]
TOTAL PLANT INVESTMENT (TPI)
STARTUP COSTS [STAR = 0.20 AOC]
INTEREST ON CONSTRUCTION CAPITAL 4
TOTAL CAPITAL REQUIREMENT (TCR)

1. SEE DEFINITION ON PAGE 58.
2. COST WOULD NORMALLY BE INCLUDED ONLY IF PURCHASE IS REQUIRED. COST IS USUALLY SMALL AND HAS NOT BEEN INCLUDED IN MODEL.
3. SEE NOTE 3 OF FIGURE 4.2.
4. SEE FIGURE 4.2.

FIGURE 4.2

RELATIONSHIP BETWEEN PRODUCTION COST FACTORS IN THE GENERAL COST MODEL


RAW MATERIALS
UTILITIES
CATALYSTS & CHEMICALS
WASTE DISPOSAL
BY-PRODUCT CREDIT
COST OF MATERIALS LESS BY-PRODUCT CREDITS (ANR)
OPERATING LABOR & SUPERVISION (AOL)
MAINTENANCE LABOR & MATERIALS [AML = 0.018 TPI]
PLANT SUPPLIES & REPLACEMENTS [APS = 0.02 TPI]
ADMINISTRATIVE & PLANT OVERHEADS [AOH = 0.70 (AOL + AML)]
DEPRECIATION [ACR = (TCR-WKC)/YEARS]
COST OF MONEY [AIC = 0.054 TCR]
1
FEDERAL INCOME TAX [AFT = 0.018 TCR]
2
LOCAL TAX & INSURANCE [ATI = 0.027 TPI]
DIRECT & INDIRECT COST
FIXED COST
TOTAL ANNUAL PRODUCTION COST [TAC]

1. AVERAGE OVER THE PLANT LIFE, ASSUMING 75% DEBT AT 9% INTEREST RATE PER YEAR, AND 25% EQUITY GIVING A NET RETURN OF 15%.
2. AVERAGE OVER THE PLANT LIFE, ASSUMING 48% FEDERAL INCOME TAX RATE.
3. ANNUAL OPERATING COST IS: AOC = ANR + AOL + AML + APS + AOH + ATI.

FIGURE 4.3

LOCATION FACTORS FOR SELECTED CITIES

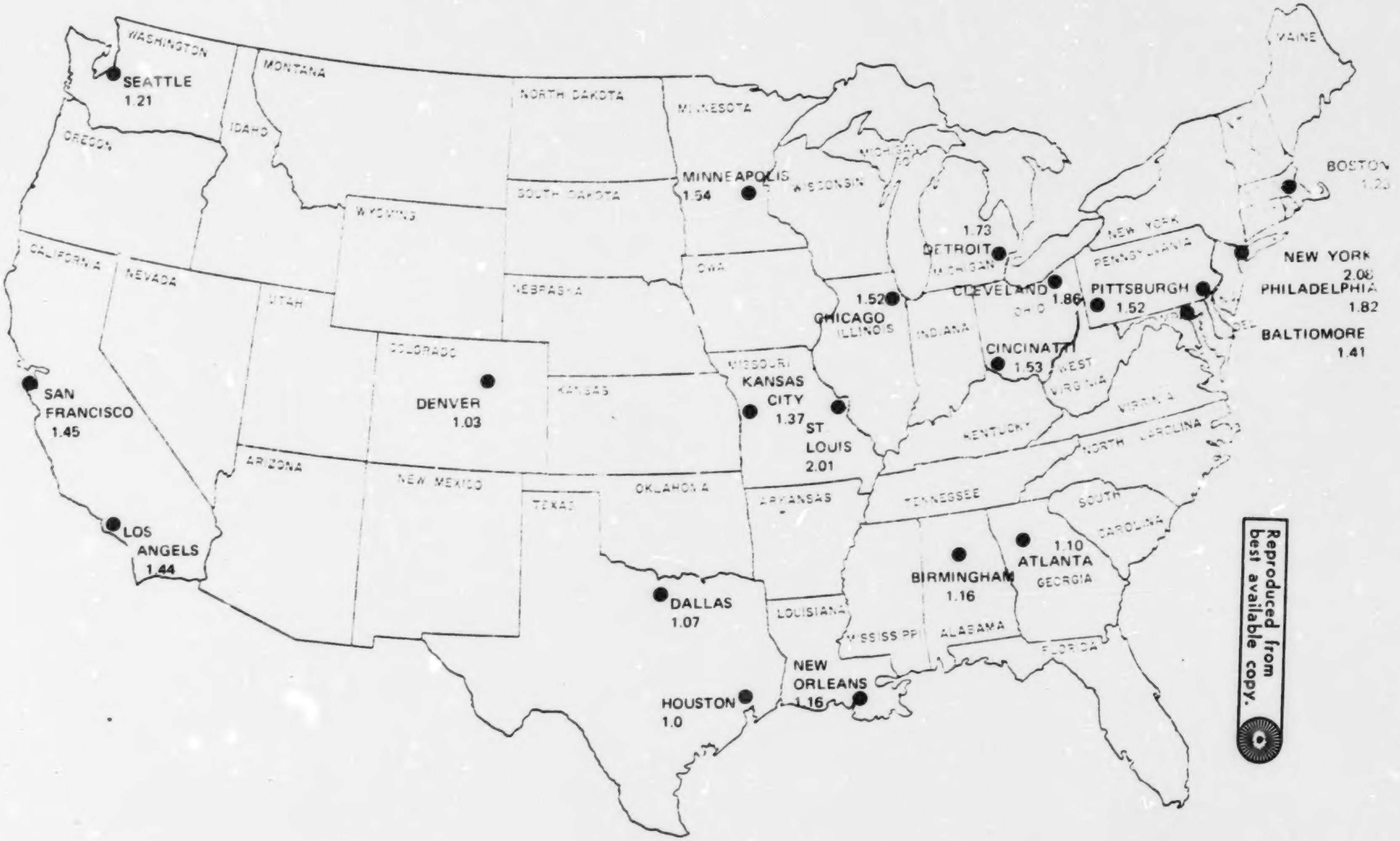

SEATTLE 1.21
SAN FRANCISCO 1.45
LOS ANGELS 1.44
DENVER 1.03
DALLAS 1.07
HOUSTON 1.0
MINNEAPOLIS 1.54
KANSAS CITY 1.37
ST LOUIS 2.01
NEW ORLEANS 1.16
CHICAGO 1.52
DETROIT 1.73
CLEVELAND 1.86
CINCINATTI 1.53
PITTSBURGH 1.52
BIRMINGHAM 1.16
ATLANTA 1.10
BOSTON 1.25
NEW YORK 2.08
PHILADELPHIA 1.82
BALTIMORE 1.41
WASHINGTON
MONTANA
OREGON
IDAHO
NORTH DAKOTA
MINNESOTA
SOUTH DAKOTA
WYOMING
WISCONSIN
MICHIGAN
CALIFORNIA
NEVADA
UTAH
NEBRASKA
IOWA
ILLINOIS
INDIANA
OHIO
COLORADO
KANSAS
MISSOURI
WEST VIRGINIA
VIRGINIA
KENTUCKY
NORTH CAROLINA
ARIZONA
NEW MEXICO
TEXAS
OKLAHOMA
ARKANSAS
TENNESSEE
SOUTH CAROLINA
GEORGIA
ALABAMA
MISSISSIPPI
LOUISIANA
FLORIDA
NEW YORK
PENNSYLVANIA
MAINE
DEL

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FIGURE 4.4

AVERAGE LOCATION FACTORS BY STATE

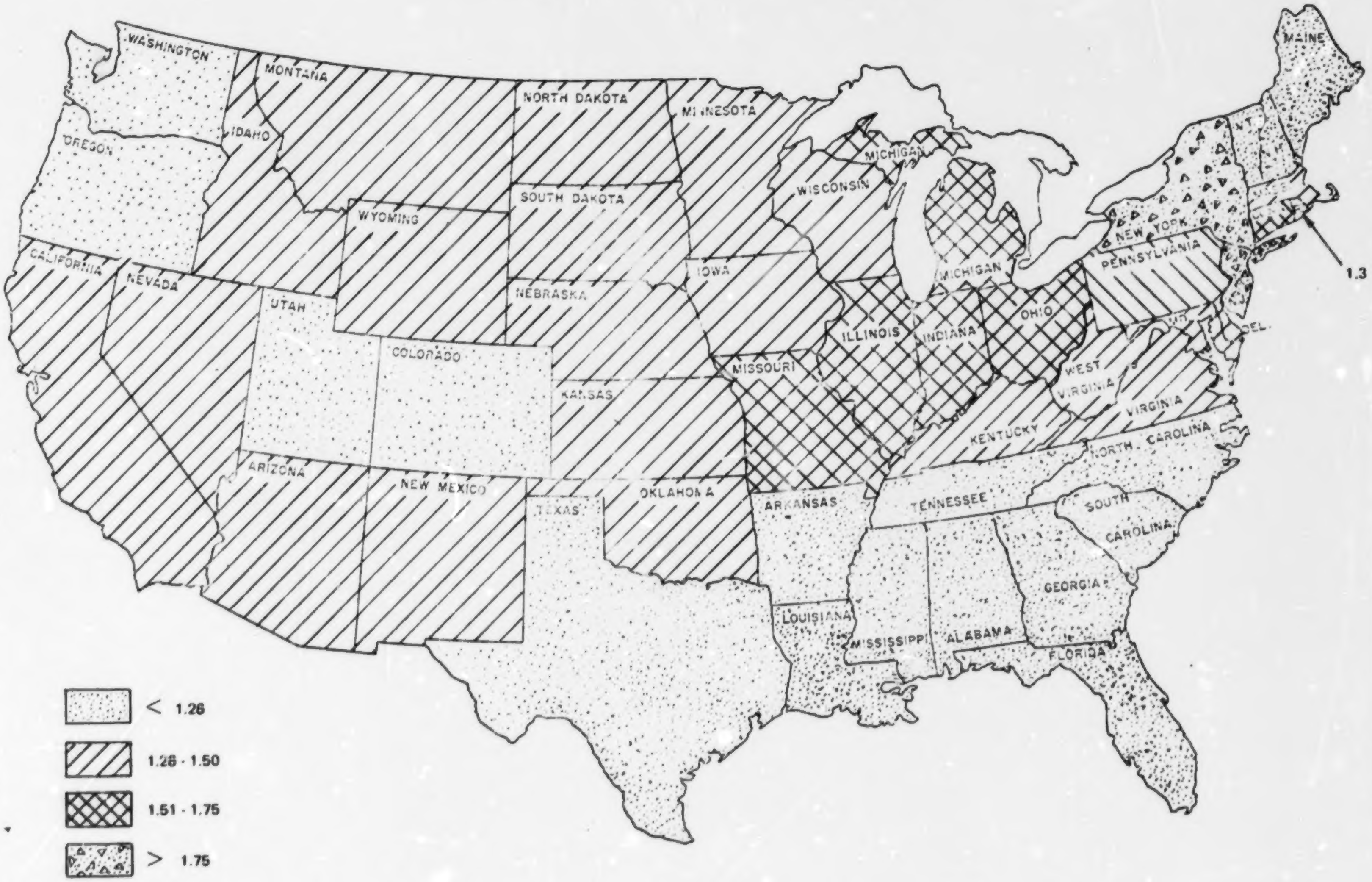
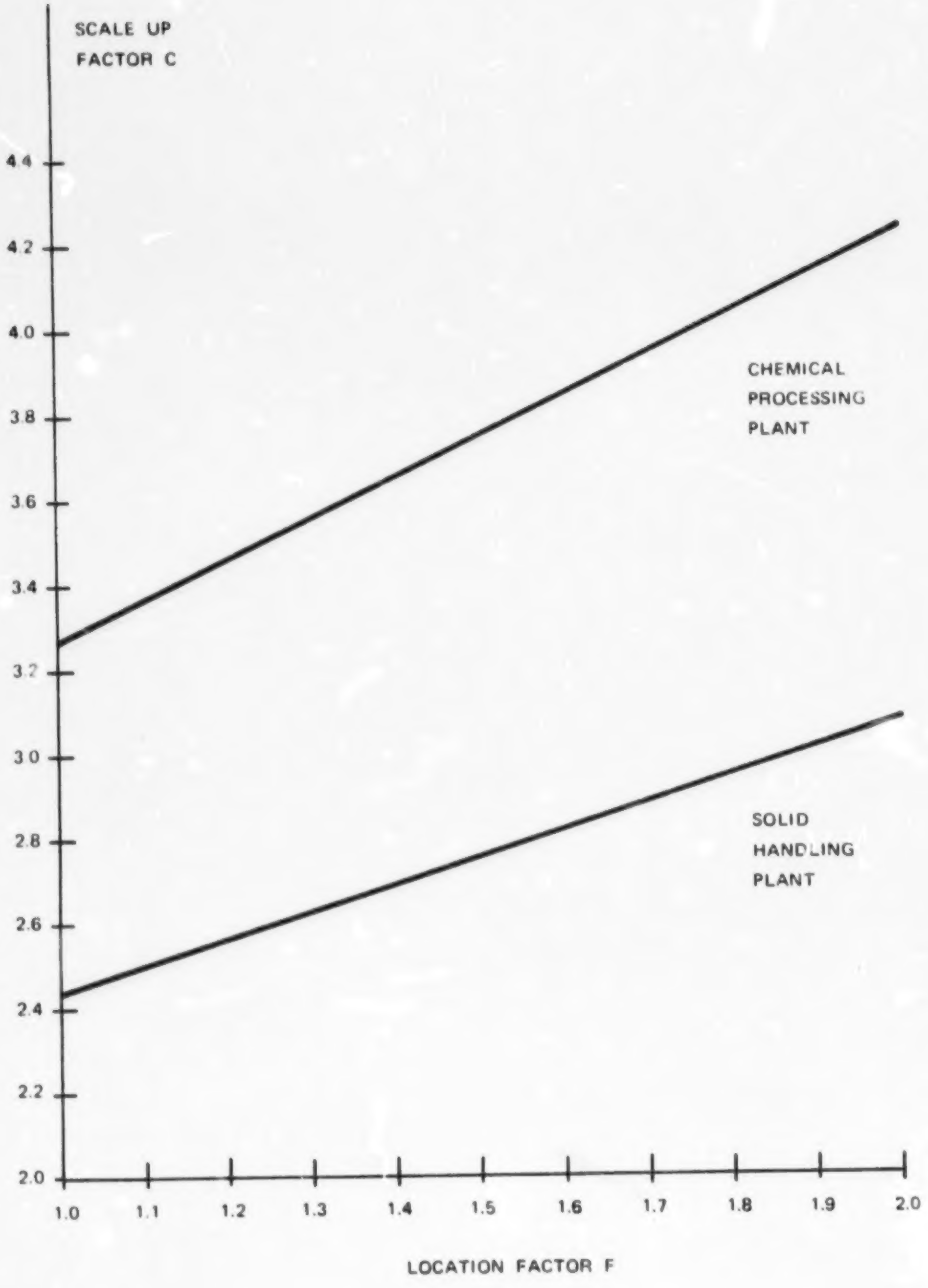

WASHINGTON
OREGON
CALIFORNIA
IDAHO
NEVADA
MONTANA
WYOMING
UTAH
ARIZONA
COLORADO
NEW MEXICO
NORTH DAKOTA
SOUTH DAKOTA
NEBRASKA
KANSAS
OKLAHOMA
TEXAS
MINNESOTA
IOWA
MISSOURI
ARKANSAS
LOUISIANA
WISCONSIN
MICHIGAN
ILLINOIS
INDIANA
MICHIGAN
OHIO
KENTUCKY
TENNESSEE
MISSISSIPPI
ALABAMA
GEORGIA
FLORIDA
SOUTH CAROLINA
NORTH CAROLINA
VIRGINIA
WEST VIRGINIA
PENNSYLVANIA
NEW YORK
DEL.
MAINE
VT
1.3
< 1.26
1.26 - 1.50
1.51 - 1.75
> 1.75

FIGURE 4.5

EFFECT OF LOCATION FACTOR ON TOTAL PLANT INVESTMENT (CONTINGENCY = 0)

TPI = C · E



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